RETROFITTING ANALYSIS OF INTEGRATED BIO-REFINERIES

A Thesis

by

BENJAMIN R. CORMIER

Submitted to the Office of Graduate Studies of
Texas A&M University
in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

December 2005

Major Subject: Chemical Engineering

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Approved by:

Chair of Committee,	Mahmoud M. El-Halwagi
Committee Members,	Mark Holtzapple
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ABSTRACT

Retrofitting Analysis of Integrated Bio-Refineries. (December 2005)

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A bio-refinery is a processing facility that produces liquid transportation fuels and/or value-added chemicals and other products. Because of the dwindling resources and escalating prices of fossil fuels, there are emerging situations in which the economic performance of fossil-based facilities can be enhanced by retrofitting and incorporation of bio-mass feedstocks. These systems can be regarded as bio-refineries or integrated fossil-bio-refineries. This work presents a retrofitting analysis to integrated bio-refineries. Focus is given to the problem of process modification to an existing plant by considering capacity expansion and material substitution with biomass feedstocks. Process integration studies were conducted to determine cost-effective strategies for enhancing production and for incorporating biomass into the process. Energy and mass integration approaches were used to induce synergism and to reduce cost by exchanging heat, material utilities, and by sharing equipment. Cost-benefit analysis was used to guide the decision-making process and to compare various production routes. Ethanol production from two routes was used as a case study to illustrate the applicability of the proposed approach and the results were bio-refinery has become more attractive then fossil-refinery.

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CHAPTER I

INTRODUCTION

Industrial processes are currently facing many challenges. Market conditions are forcing changes in quantities and qualities of various products. With the qualitative and quantitative changes, production technologies and feedstocks must be re-assessed. Additionally, existing capital investments must be optimally utilized to enhance capital productivity. In some cases, an existing facility may be integrated with a new production line. Feedstocks may be substituted or mixed with other raw materials. In such cases, the economics of the existing plant can be improved when the process is combined with the new plant. Economy of scale, technological upgrades, debottlenecking activities, and heat/mass integration between the two facilities can provide cost savings and reduction in energy and material utilities.

One of the emerging concepts in the process industry is the idea of a bio-refinery. One way of defining a bio-refinery is that it is a facility which converts biomass into liquid transportation fuels or valuable chemicals and other products. Because of the renewable nature of biomass, a bio-refinery is an important element in a circular economy which seeks to minimize negative ecological impact. The biomass used in bio-refineries may be in different forms such as forest products, agricultural products, and bio-wastes. Figure 1 is a schematic representation of a bio-refinery with its feedstocks and products.

This thesis follows the style and format of Clean Technologies and Environmental Policy.

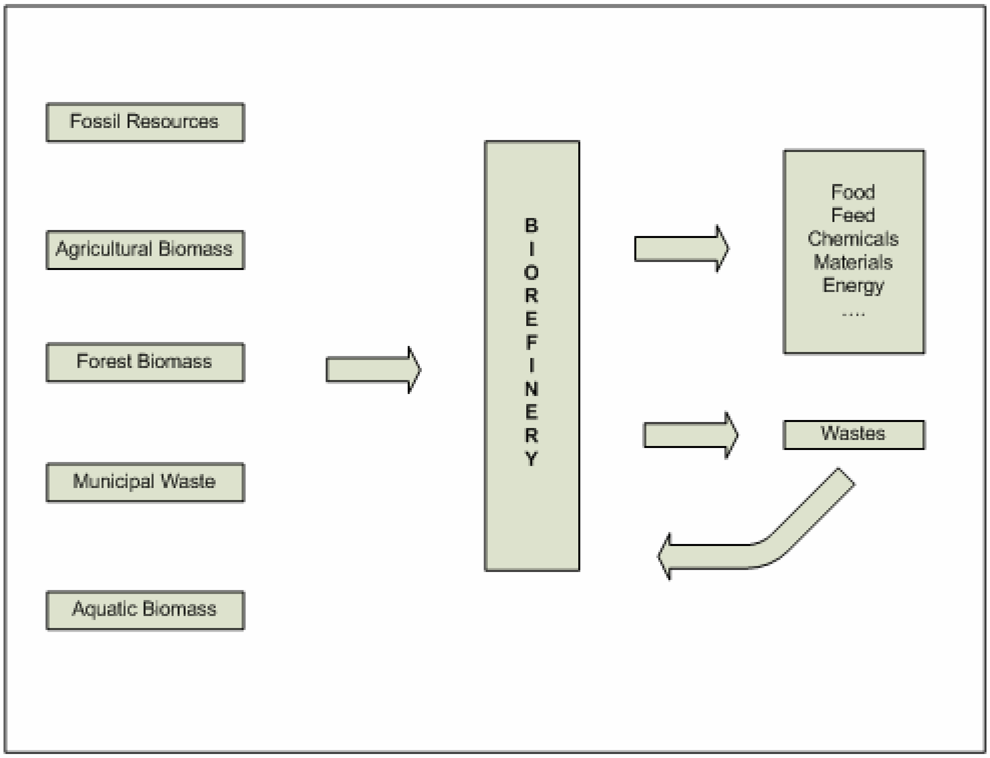

Figure 1.1 Bio-refinery concept

Over the past decade, scientists have raised concern about the sustainability of the traditional industries and how some practices may jeopardize our future. Depletion of natural resources and discharge of hazardous wastes are disconcerting factors. The greenhouse gas emission has also become a cause of concern (Mintzer and Schartz 2003). A bio-refinery can address some of these concerns because it involves the usage of renewable resources, it is part of a sustainable life cycle, and it normally results in less green house gas emissions compared to fossil-based processes. As such, a bio-refinery is regarded as a serious alternative to produce energy and chemicals in the near future.

In light of the dwindling resources of fossil fuels, the following questions must be addressed:

- What will happen to the existing refineries and to the enormous capital investment of these refineries?
- Can biomass be used to supplement or substitute the fossil-fuel feedstocks?
- Can bio-refineries be integrated with the traditional refineries and chemical-producing facilities?
- What would be a transitional strategy to move from traditional plants to bio refineries?
- Under what conditions will bio-refineries be competitive in the US?

To begin to address these issues, it is necessary to study the retrofitting of existing facilities by integrating them with bio-refineries and/or by substituting their feedstocks with biomass. This work is aimed at developing a retrofitting analysis of integrated bio-refineries. In particular, the work will study the prospects of using heat and mass integration to reduce energy and material utilities and to enhance to economic performance of the original facility and the bio-refinery.

CHAPTER II

BACKGROUND AND LITERATURE SURVEY

In the 1970's, there was an initial surge of interest in biomass-for-energy facilities following the energy crisis and the increase in crude oil prices. Many of the original ideas were not commercially pursued because of the subsequent decline in oil prices. However, recently the situation has changed with the increasing attention to global warming and the increasing prices of fossil fuels. The biomass-based fuels and products can make contributions to the U.S. environment, chemicals, power, and economy. As such, the concept of a biorefinery seems to be a serious option to improve the condition of our future while enhancing our economy.

In this chapter, the background and literature survey will be presented in the following order: the current definition of biorefinery, characterization and availability of biomass feedstock, economic aspect, and chemical available through biorefinery.

2.1 Biorefinery

Future refineries will encounter feedstock problems in terms of availability and cost. As the price of crude oil increases and the supplies diminish, there will be a need to find alternative sources of energy feedstocks. One solution of these issues is called biorefinery. The term biorefinery was recently coined to address refineries capable of converting biomass to valuable chemicals or energy. As more renewable sources are used, less waste is generated in the ecological cycle.

Biomass products can range from biomaterials to fuels or important feedsocks for the productions of chemicals and other materials. Biorefineries can be based on a number of processing platforms using mechanical, thermal, chemical, and biochemical processes. Biorefinery conversion of biomass should be considered not only as a substitute for fossil resources but also as an integrated use of living organisms, microorganisms and enzymes in a cycle of producing foodstuff, fuels, feeds, value-added chemicals and industrial materials. In the literature (Sophonsiri and Morgenroth 2004), there are numerous examples showing that municipal wastes can be decomposed into valuable products such as energy, chemicals, and byproducts.

2.2 Characterization and availability of biomass feedstocks

The biomass feedstock is divided in two categories, grains and lignocellulose. Grains feedstock is such as corn and other corps. Lignocellulose feedstock is composed of cellulose and hemi-cellulose, which represent about 80% of biomass feedstock. The different types of feedstock are shown in Table 2.1. Herbaceous and woody biomass are composed of carbohydrate polymers such as cellulose and hemicellulose, lignin and small parts of acids, salts and minerals. The cellulose and hemicellulose compose about two-thirds of a dry biomass.

The following is a typical composition of lignocellulose (Scurlock 2003):

- *Cellulose*, between 40% and 60% of the dry biomass, is formed of glucose-glucose dimer. Hydrolysis is needed to break down the hydrogen bond (hard to break) and the product, glucose, is a six-carbon sugar.
- *Hemicellulose*, between 20% and 40% of dry biomass, consist of short highly branched chains of various sugars, mainly five-carbon and six-carbon as xylose, arabinose, galactose, glucose and mannose. Hemicelluloses are easy to break down during the hydrolysis.

- *Lignin*, between 10% and 25% of dry biomass, counted as a residue during the ethanol process.

Table 2.1 Different feedstocks possible for lignocellulose process

Herbaceous	Woody	
	Hardwood	Softwood
Agriculture residues Corn stover Sugar cane bagas Wheat strove Rice strove Rice hulls Energy corps Switch grass Energy can Sweet sorghum Bana grass	Hybrid popular wood Aspen wood	Logging residues

The key element for biorefinery is the availability of biomass. According to a recent study by the Department of Energy, the current quantity of biomass in the entire U.S. is about half of a billion dry tons (U.S. Department of Agriculture 2005). This study has shown the different categories of biomass and their quantities. These quantities are summarized by Table 2.2.

Table 2.2 Current availability of biomass feedstocks (U.S. Department of Agriculture 2005)

Biomass from forest resources	Quantity (million dry tons)
Logging activities, cultural operations and clearing of timberlands	67
Fuel treatment operations on timberland and forestland	60
Roundwood to energy (fuelwood)	35
Industry and urban wood wastes	36-106
Forest growth and increase in the demand will generate biomass	89
Biomass from agricultural lands	**Quantity (million dry tons)**
MSW & other wastes	25
Manures	35
Grains (biofuels)	18
Other crop residues	21
Small grain residues	7
Wheat straw	13
Corn stover	75

The study by the Department of Energy also predicts that over one billion ton of biomass would be available after some modifications of the biomass. These potential quantities are summarized by Table 2. 3.

Table 2.3 Potential availability of biomass under increased crop yields and technology changes (U.S. Department of Agriculture and U.S. Department of Energy 2005)

Type	Quantity (million dry tons)
MSW & other wastes	31
Manures	44
Grains (bio-fuels)	55-95
Other crop residues	37-49
Soybeans	26
Small grain residues	19-32
Wheat straw	42-72
Corn stover	152-230
Perennial crops	156-377

The forestland and agricultural land are the two largest potential biomass sources. Currently 1.3 billions dry ton per year is available, this can supply one-third of the actually demand of transport fuels. Biomass supplies about 3% of energy consumption in the United State from the pulp and paper industry and electric generation using forest industry residues and municipal solid waste. Table 2.4 summarizes some of these quantities.

Table 2.4 Energy and quantity representation of the biomass by categories (U.S. Department of Agriculture and U.S. Department of Energy 2005)

Type	Quantity in Millions dry tons	Quads
Urban wood wastes	36.8	0.63
Primary mill wastes	90.5	1.5
Forest residues	45	0.76
Agricultural residues	150.7	2.3
Energy crops	188	0.29
Total	511	5.48

It is estimated that biomass will supply 5% of the nation's power, 20% of its transportation fuels, and 25% of its chemicals by 2030 (U.S. Department of Agriculture and U.S. Department of Energy 2005). The surface area of agricultural land is 455 million acres, of which 349 million acres of land are in active use, 39 million acres are idle, and 67 million acres of cropland pasture over 48 states (U.S. Department of Agriculture and U.S. Department of Energy 2005). The biomass is not similar to the petroleum feedstock. Since the biomass is renewable resource.

2.3 Economics aspect of biomass feedstocks

The other side of the biomass feasibility is the price at which biomass is available. The delivered price varies as a function of many factors including the type of biomass, harvesting and collection techniques, storage, the hauling distance, the region, and the available quantity. The other side of the economic aspects of biomass utilization is the potential credit and/or subsidy associated with the reduction in greenhouse gas (GHG) emissions.

2.3.1 Delivered cost of biomass

Biomass feedstocks offer a distinct advantage being a renewable resource. For instance, every year corn has can be planted, grown, harvested and transported to the biorefinery to produce ethanol or other energy-related chemicals. The delivered cost of biomass depends on the expenses associated with growth, harvesting, and transportation. Walsh et al. (1999) published a report estimating the quantities and delivered cost of biomass for every state in the U.S. Figure 2.1 shows the distribution of the biomass available at different delivered price.

2.3.2 Greenhouse emission effect

Renewable energy such as bio-derived ethanol is produced from plants that use solar energy to grow. Combustion of a bio-fuel releases a portion of this energy. As Fig. 2. 2 shows, solar energy and the use of carbon dioxide are important steps in the life-cycle of an energy crop. This cycle entails the net release of a much smaller quantities of greenhouse gases (GHGs) compared to fossil fuels. Most of GHG emissions from process industries are related to combustion of fossil fuel. In U.S. several state are working on the legislations of GHG emissions. Therefore, bio-fuels represent one of the most promising options for reducing GHG emissions from the transportation sector.

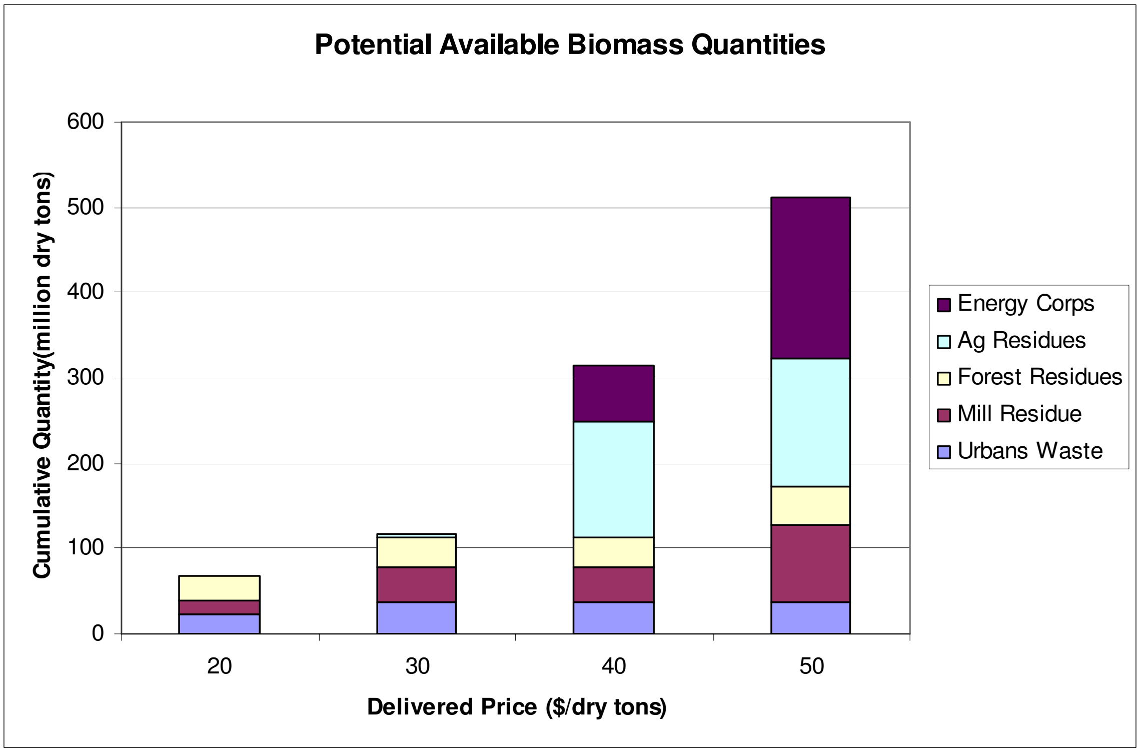

Figure 2.1 Potential available biomass quantities at different delivered price (Walsh et al., 1999)

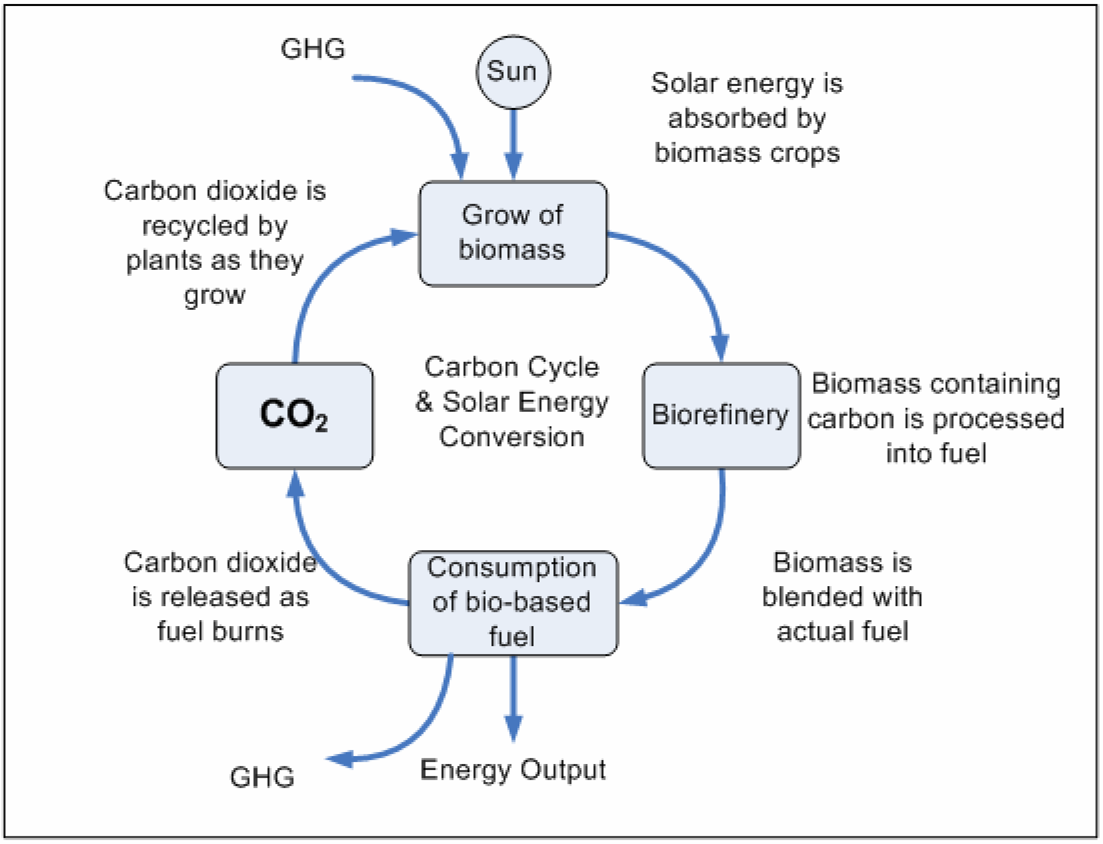

Figure 2.2 Idealized life cycle of an energy crop

2.4 Chemical potential of biomass feedstocks

Today's biomass uses include ethanol, bio-diesel, biomass power and industrial process energy. In preparing biomass for bio-refineries, there are two primary pathways: hydrolysis of cellulosic biomass to sugars and lignin, and thermo-chemical conversion of biomass to synthesis gas. Figure 2.3 is a schematic representation of these two pathways from biomass to bio-refineries.

The sugar platform is composed of thermo-chemical pretreatment and enzymatic hydrolysis. The pretreatment is usually done with dilute acid and breaks down the hemicellulose into sugars such as xylose Cellulose are enzymatically hydrolyzed to release glucose. Then, the produced sugar is available for fermentation to ethanol. Some forms of used biomass involve corn stover, sorghum and seeds.

Thermo-chemical technologies utilize catalysis and/or high pressure and temperature to convert biomass. Oils and bio-products from wood resources are used as such. The lignocellulosic biomass represents an energy potential. Gasification and pyrolysis are used to convert biomass into an energy fuel. It mostly used to produce electrical energy by cogeneration. Example of used biomass includes switch grass and wood.

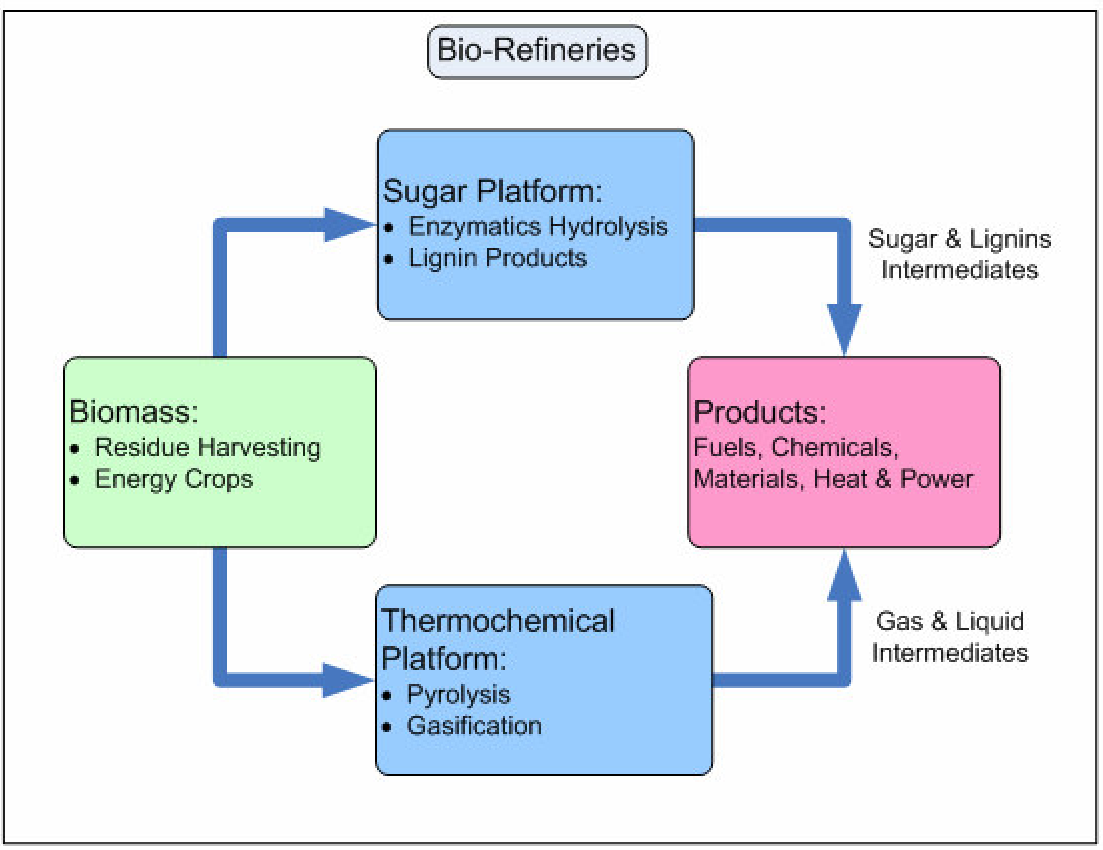

Figure 2.3 Diagram for biomass for purpose use (U.S. Department of Energy 2004)

There are also other platforms such as biogas, carbon-rich chains, plant products and bio-oil which are beyond the scope of this work. Biogas platform is the decomposition of biomass into methane and carbon dioxide by anaerobic digesters (El-Halwagi 1986) Carbon-rich chains platform is the transesterification of vegetable oil or animal fat into fatty acid methyl ester, commonly known as bio-diesel. The plant products platform is the use of selective breeding and genetic engineering that can produce greater amounts of desirable feedstock and chemicals. Bio-oil may be produced through the pyrolysis of biomass to produce oil with similar characteristics to petroleum cuts. The Table 2.5 presents the summary of different technology platforms and a sample of the chemical products.

Table 2.5 Summary of industrial bio-based product (Paster et al. 2003)

Technology Platform	Chemicals
Sugars fermentation	Lactic Acid, Polyactide, Ethyl Lactate, 1,3-propanediol, Succinic Acid & derivatives, bionolle 4,4 polyester,3-Hydorxypropionic Acid & derivatives, N-Butanol, Itaconic Acid
Sugars fermentation & Thermochemical	Propylene Glycol
Sugars thermochemical	Isosorbide, Levulinic Acid & derivatives
Oils & Lipids	Lubricants and Hydraulic Fluids, Solvents, Polymers (polyurethanes), Proteins
Biomass Gasification	Fischer-Tropsch & Gas-to-liquids Products
Biomass Pyrolysis	Phenol-Formaldehyde Resins
Biocomposites	Biocomposites
Plants as Factories	Guayule, Polyhydroxylalkanoates (PHAs)
Photosynthesis, Anaerobic Digestion	Lignin, Methane, Carbon dioxide, other chemicals.

From these chemical, the potential derivatives of these chemicals is endless. For instance, it is possible to derive eight other chemicals from lactic acid and so and so. These chemicals represent an attractive target for new bio-based products. Table 2.6 represents the current bio-products for different biomass with their different technology required.

Table 2.6 Current industrial bio-products form biomass (Paster et al. 2003)

Category	Principal Technologies	Feedstock	Chemical
Starch & Sugars	Biochemical	Biomass sugars derived from corn and sorghum	Lactic acid, citric acid, ethanol, starch, sorbitol, levulinic acid, itaconic acid
Oil & Lipids	Thermochemical	Oils/lipids derived from soybean rapeseed	Glycerol/glycerine, alkyd resins, high erucic acid rapeseed, polyurethane, epoxidized soybean oil, factice, sulfurized fatty oils, fatty acids, cyclopentadienized oils, lecithin, maleinized oils
Specialty Crops	Thermochemical	Spearmint, peppermint, sweet almond	Spearmint oil, peppermint oil, sweet almond oil
Forest derivatives	Thermochemical	Pine, black liquor and soft wood	Turpentine oil, rosin, tall oil, and cellulose derivatives(esters, acetates, etc)

Waste is also considered as part of biomass, the municipal, industrial and agriculture wastes can be broken down into simple compounds such as proteins, carbohydrates, and lipids (Sophonsiri and Morgenroth 2004). A Bio-refinery may produce a wide variety of potential chemicals

One of the most promising bio-refinery products is ethanol. Ethanol is mainly produced by fermentation (95% fermentation, 5% synthetic from ethylene). Ethanol could be used as an alcoholic beverage, industrial alcohol, and/or fuel-related alcohol

(Berg 2003). Ethanol represents an easy bio-fuels alternative for the near future. Ethanol is less expensive than the other oxygenates, octane enhancers and often conventional gasoline. Ethanol is also seen as a way of reducing dependence on foreign sources of oil and gas. It can also lead to enhancing environmental quality. Ethanol reduces vehicle emissions. For instance, according to Dinneen (2005), if a10%-ethanol 90%-gasoline fuel blendis used, this will:

- Reduce tailpipe fine particulate matter (PM) emissions by 50%
- Reduce secondary PM formation by diluting aromatic content in gasoline
- Reduce carbon monoxide (CO) emissions by up to 30%
- Reduce toxics content by 13% (mass)

The United States is not the only country that is closely considering expanding the use of ethanol as a bio-fuel. Table 2.7 illustrates the worldwide production of ethanol.

Table 2.7 World ethanol production (all grades, in millions of gallons) (Dinneen 2005)

Brazil	3989	Italy	40
U.S.	3535	Australia	33
China	964	Japan	31
India	462	Pakistan	26
France	219	Sweden	26
Russia	198	Philippines	26
South Africa	110	South Korea	22
U.K.	106	Guatemala	17
Saudi Arabia	79	Cuba	16
Spain	79	Ecuador	12
Thailand	74	Mexico	9
Germany	71	Nicaragua	8
Ukraine	66	Mauritius	6
Canada	61	Zimbabwe	6
Poland	53	Kenya	3
Indonesia	42	Swaziland	3
Argentina	42	Others	338
		Total	10770

Many countries try to reduce petroleum imports, enhance the air conditions, and boost their agricultural economic. Ethanol addresses all of these objectives. The growth of ethanol production in these countries is estimated to accelerate as the countries account for the greenhouse gas emissions. The United States has witnesses an increasing level of interest in ethanol production the growth of the production is shown in Fig. 2.4

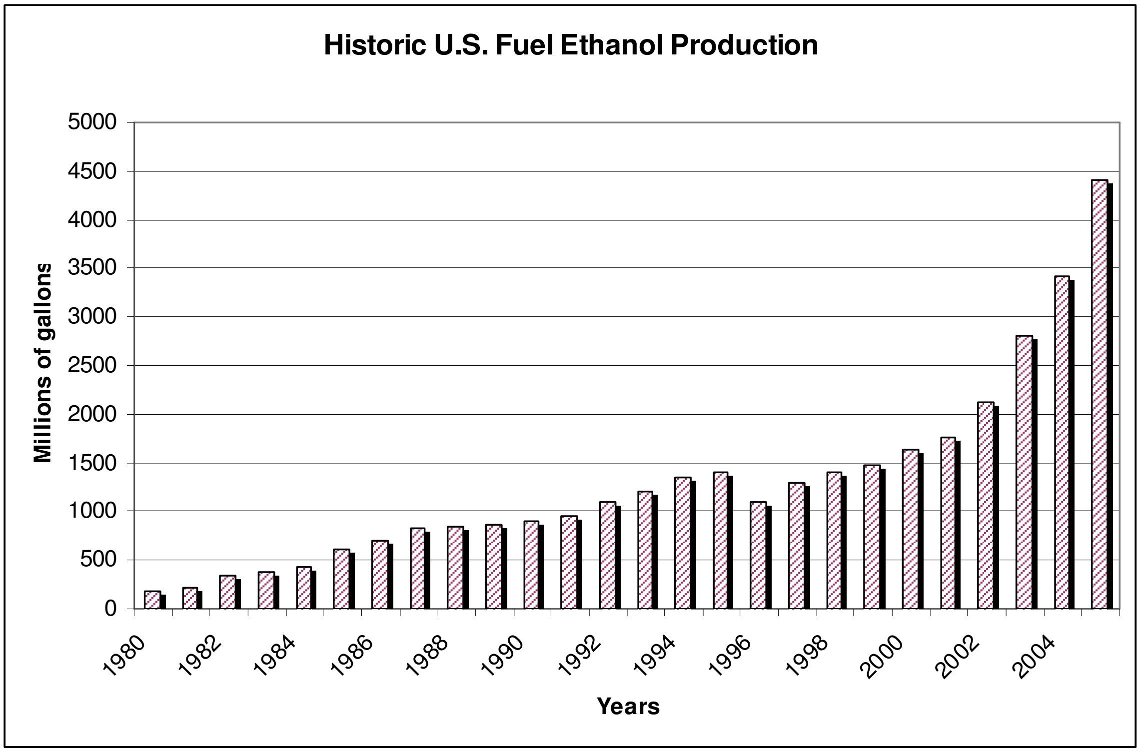

Figure 2.4 Historic U.S. fuel ethanol production (Dinneen 2005)

For 2005, the production of ethanol is estimated at 4,400 millions of gallons. It is expected that the U.S. will become the top producer of ethanol surpassing Brazil which has remained the world's leafing producer of ethanol for the past half a century The U.S. ethanol plants are mainly situated in 20 states throughout the country. The biggest plants are in the north of the U.S. in states such as Iowa, Illinois, Nebraska, and Minnesota.

The fermentation process is the dominant process in the market (95% of ethanol production). Ethanol is fermented from grain and some from lignocellulose.

Grain-derived ethanol is produced via the wet-milling or dry-milling processes (BBI International 2003). In Fig. 2.5, both processes are described. In the dry-mill process, the corn is ground into flour and is processed without any separation of component parts. In the wet-mill process, the corn is soaked or steeped then separated into its component parts, which are recovered prior to fermentation.

Both processes release carbon dioxide while producing ethanol. The key difference is in the set of byproducts. The demand for these byproducts is an important factor determining the type of the plant.

Notwithstanding several advantages of bio-derived ethanol, these plants have limitations. These limitations may include the dependence on natural gas, the geographical position, the presence of sufficient market demand for the byproducts, competition and availability for feedstock, and the sensitivity of fermentation. When these limitations are relevant, they must be addressed and balanced against the benefits of bio-derived ethanol.

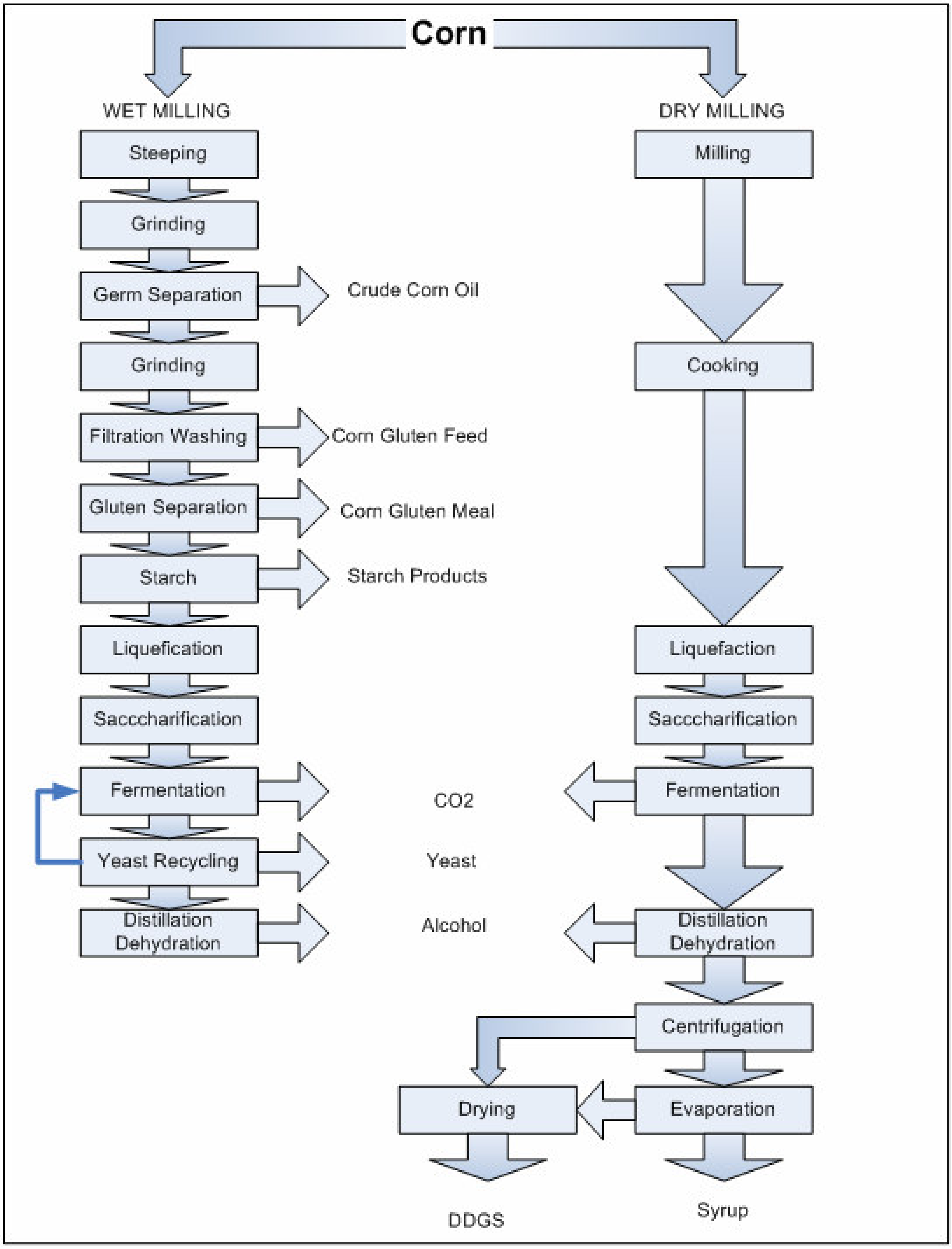

Figure 2.5 Corn-based process flow diagrams (BBI International 2003)

CHAPTER III

PROBLEM MOTIVATION, STATEMENT, AND CHALLENGES

3.1 Motivation

There is an escalating need for chemical and energy products that may be derived from biomass and/or fossil sources. To respond to these needs, it is necessary to retrofit existing facilities or add new production lines (e.g., bio-refineries). Process integration provides an attractive framework for retrofitting existing facilities or integrating new facilities with neighboring plants. Integration may take several forms including:

- Equipment sharing
- Feedstock allocation and/or substitution
- Energy integration (e.g., heat integration through the synthesis of a heat-exchange network, common utility "islands", etc.)
- Mass integration (e.g., material reuse, separation network synthesis, etc.)
- Waste handling (e.g., common treatment facilities, waste exchange, etc.)

While philosophically it is sensible to adopt integration in retrofitting and/or development of a new facility, the questions is how? There is a need to develop a systematic and generally applicable design procedure which guides process engineers as they make their decisions in retrofitting and integration. Current literature lacks such a procedure and the current work is aimed at addressing this research gap.

3.2 Problem statement

Given an existing plant, which uses F_{fossil} fossil-bases feedstock and produces $P^{existing}$ flowrate of product, it is desired to increase product flowrate to $P^{retrofitted}$. The product may be the same as that produced by the fossil-based facility or may be different (in terms of quality or type). The following degrees of freedom are available:

- Material co-feeding or substitution: Biomass feedstock, $F_{biomass}$ (in addition to or in lieu of fossil-based feedstock)
- Unit adaptation: Some existing units may be used for processing bio-based feedstock, $U_{existing}$
- Unit addition: New units may be added to retrofit the plant, U_{new}
- Process integration: Integration of mass and energy utilities ($M_{utilities}$, and $E_{utilities}$, respectively) among various parts of the process (existing, added etc.),

A schematic representation of the stated problem is given by Fig. 3.1. The stated problem is a general bio-refinery retrofitting problem which involves the integration of a bio-refinery with a traditional processing facility. To focus the discussion, a specific problem will be used as a case study. This problem is the production of ethanol which is among the most promising products from bio-refineries. Consider a current fossil-based plant (ethylene feedstock) which produces ethanol. Because of the anticipated increase in ethanol demand, the plant is interested in increasing its production. To retrofit the plant, two options are considered: using the same ethylene-based technology while increasing capacity and/or adding a bio-refinery which uses fermentation of a biomass feedstock. Both production lines should be integrated. Mass and energy integration will be used to reduce energy and material utilities and to identify shared equipment. A systematic procedure is to be proposed to design the integrated facility and to assess the various options.

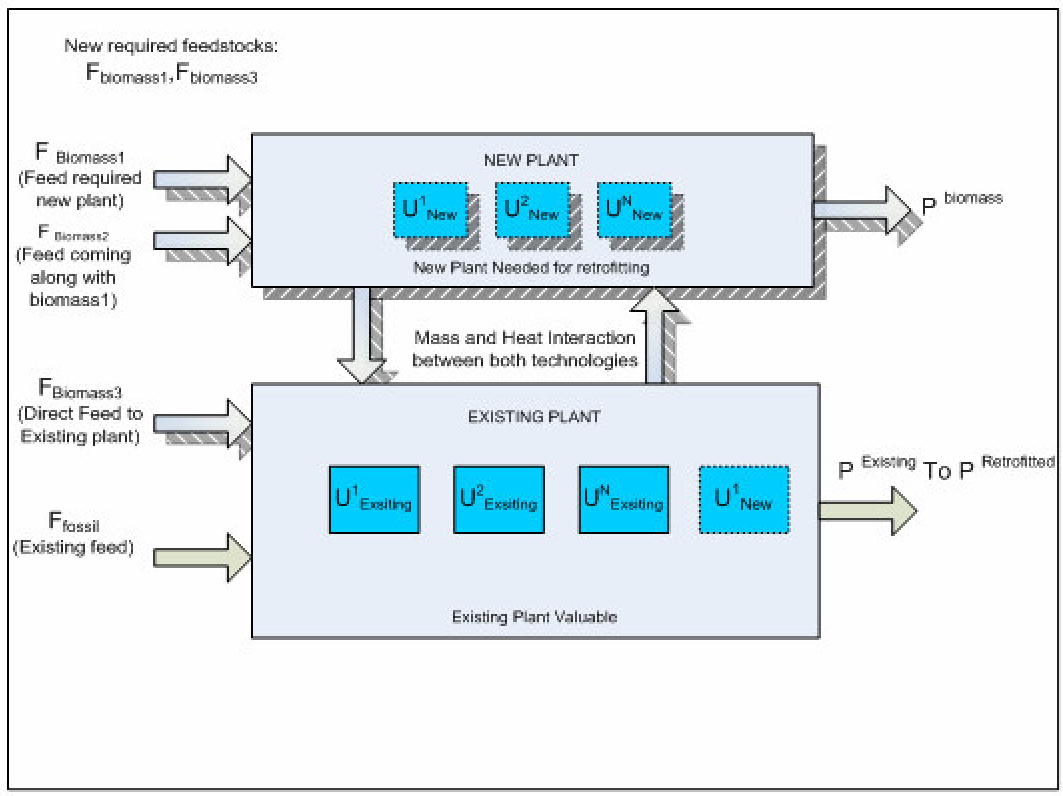

Figure 3.1 Schematic representation of the problem statement

3.3 Design challenges and objectives

The aforementioned problem involves addressing the following challenges:

- What will happen to the existing refineries and to the enormous capital investment of these refineries?
- Can biomass be used to supplement or substitute the fossil-fuel feedstocks?

- Can bio-refineries be integrated with the traditional refineries and chemical-producing facilities?
- What would be a transitional strategy to move from traditional plants to bio refineries?
- Under what conditions will bio-refineries become competitive in the US?

These challenges call for the development and application of a systematic methodology. This methodology should include tools that enable process retrofitting with the generation of alternatives, analysis, process integration, and economic evaluation. This work is aimed at providing the first steps in addressing these challenges. In particular, the specific objectives of this work are:

- Develop a hierarchical framework for systematically addressing the retrofitting and integration design tasks. This should framework decompose the problem into several manageable stages.
- Develop a systematic approach to coordinate the use of different process analysis and synthesis tools (e.g., simulation, cost estimation, synthesis of heat exchange networks, synthesis of material recycle networks, etc.).
- Demonstrate applicability and value of process integration in retrofitting and integrating bio-refineries.
- Illustrate the usefulness of the developed framework and design procedure by addressing a case study on ethanol production.

Chapter IV presents the rationale and details of the design procedure. The case study on ethanol production is presented in Chapter V. The conclusions of this work and the recommendations for future research are described by Chapter VI.

CHAPTER IV

METHODOLOGY

As stated in the problem statement, the purpose of this work is to develop a methodology to extend an existing plant either by increasing the same feedstock or by retrofitting new feedstock like biomass. Figure 4.1 shows the proposed methodology to address this problem.

This proposed design approach is a hierarchical framework. There are three building blocks that are arranged in order of increasing cost. First, effort is made to reach the production target using no or low cost strategies. These include process reconfiguration (e.g., stream rerouting) and modification of operating conditions. Next, medium-cost modifications are pursued. These include the addition of new units and/or th replacement of existing units with new ones. Finally, capital-intensive strategies are invoked. These include the addition of new production lines. Each design step involves the use of several tools for analysis, integration, and cost evaluation. These tools are described in the following sections.

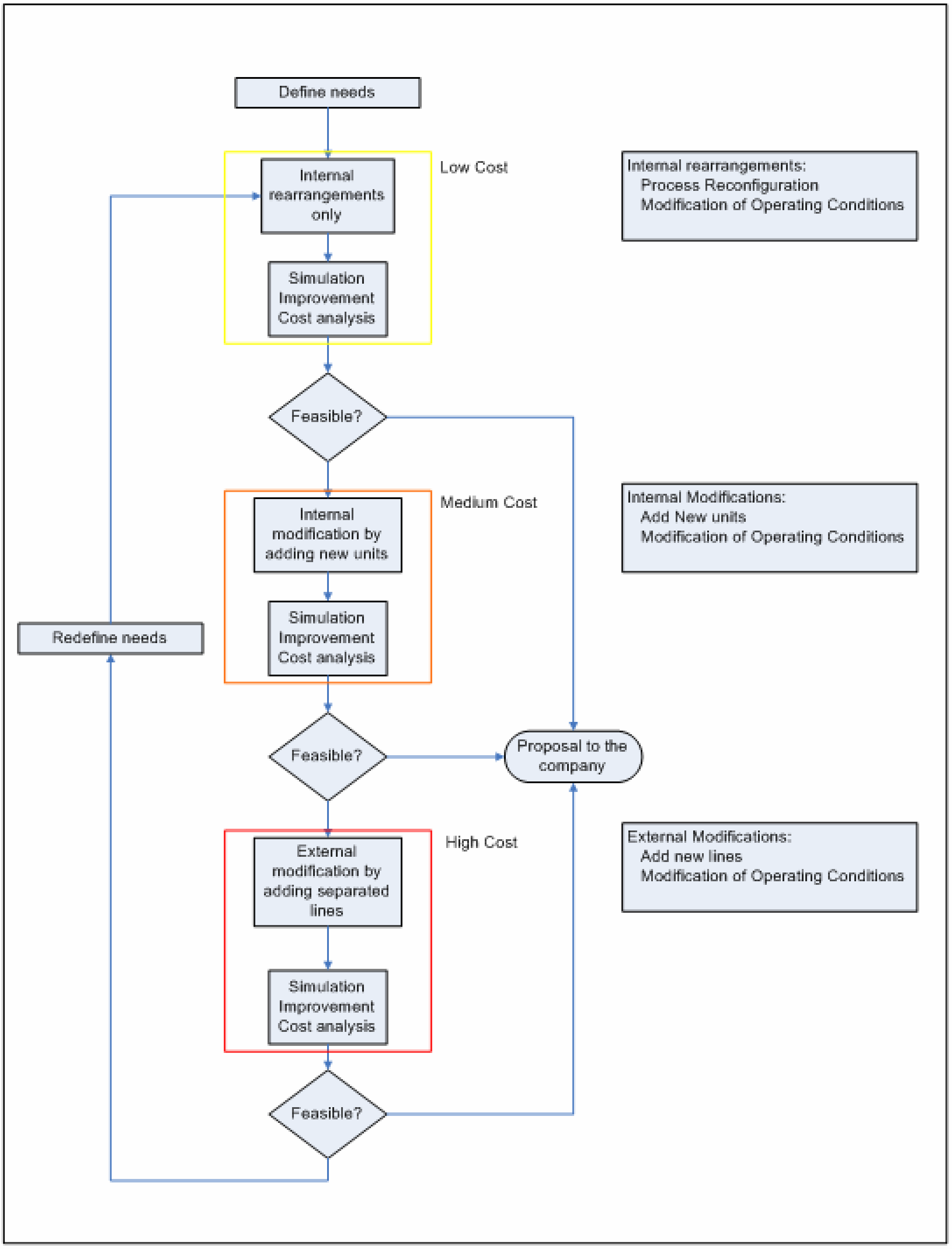

Figure 4.1 Proposed flowchart

4.1 Tools

To the following sections describe the tools to be used to optimize, simulate, and analyze the feasibility of a plant.

4.1.1 Simulation

A chemical plant can be modeled with a computer-aided simulator such as Aspen Plus© to evaluate the flowrate, reaction and thermodynamic behavior of the chemicals. Furthermore, such a model can be used to simulate any change that the company wants to do. Aspen Plus© has an extensive thermodynamic database and simulates a close model from the reality. It will simulate the feasibility of the change by using thermodynamics.

4.1.2 Process integration and optimization

There are two ways to improve the mass and heat utilities, by using material rerouting and heat exchanger network.

4.1.2.1 Material rerouting network

A general framework for material rerouting has been developed by El-Halwagi et al. (2003). In this framework, process sources (streams carrying targeted species) are rerouted to process sinks (units that can employ these sources). Constraints are imposed on the quantity and characteristics of the feed to each process sink. Strategies of stream segregation, mixing, and allocation are used to determine the optimal rerouting scheme. A graphical technique, called material recovery pinch analysis is used to identify rigorous targets for fresh usage, material recycle, and waste discharge (El-Halwagi et al. 2003). This technique can be used to identify targets for minimum fresh usage, maximum stream recycle, and minimum waste discharge. The first step is to determine what the targeted species is/are. Then for each operation, the sinks have to be ranked in ascending order of maximum admissible composition as shown in the Table 4.1.

Table 4.1 Sink distribution streams

Sink	Flow	Maximum Inlet Concentration	Load
1	F_1^{Sink}	C_1^{Sink}	$M_1^{sink,max}$
2	F_2^{Sink}	C_2^{Sink}	$M_2^{sink,max}$
…	…	…	…
J	F_j^{Sink}	C_j^{Sink}	$M_j^{sink,max}$

where

$$M_i^{Sink,Max} = F_i^{Sink} \times C_i^{Sink} \qquad (4\text{-}4)$$

The sources have to be ranked in ascending order of pollutant composition as shown in the Table 4. 2.

Table 4.2 Source distribution streams

Source	Flow	Concentration	Load
1	F_1^{Source}	C_1^{Source}	M_1^{Source}
2	F_2^{Source}	C_2^{Source}	M_2^{Source}
…	…	…	…
J	F_j^{Source}	C_j^{Source}	M_j^{Source}

Where

$$M_i^{Source} = F_i^{Source} \times C_i^{Source} \qquad (4\text{-}5)$$

Figure 4.2 represents the plot of the maximum load as a function of its flowrate. This graph has two curves, one is the sink and the other one is the source. The sink arrows are superimposed in ascending order and as well for the sources. The two composites are slid on the fresh locus until the both curves touch with the source composite completely below the sink composite. The point where they touch is the material recovery pinch point.

This technique can be used for fresh and waste minimization (e.g., minimization of fresh water and wastewater discharge) and it would have three main achievable targets:

- Analysis: Identifying the minimum fresh resource consumption and minimum waste discharge operations
- Synthesis: Designing a water-using network that achieves the identified flowrate targets for freshwater and wastewater through water reuse, regeneration and recycle
- Retrofit: modify an existing water-using network to maximize water reuse and minimize wastewater generation through effective process changes

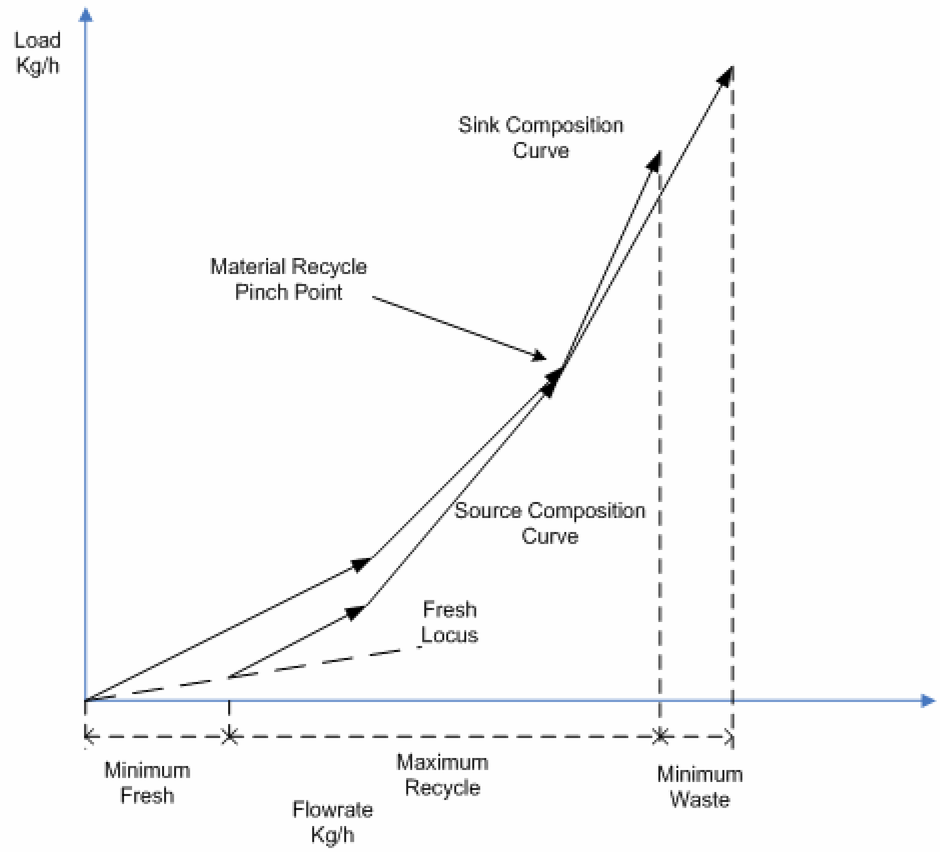

Figure 4.2 Illustration for material recovery pinch diagram (El-Halwagi et al., 2003)

4.1.2.2 Heat Exchanger Network

In the plant, heating and cooling represent an important operating cost. In order to minimize the operating cost for the heat utilities, heat integration is needed. This techniques was developed by Linnhoff and coworkers in the 90's (Ahmad et al. 1990; Linnhoff and Ahmad 1990). In order to achieve the optimal network the design, targets are needed. To systematically synthesize several optimal network of exchanger, heater and/or cooler in order to reach the desired stream output (target) temperatures and achieve the following multiple design objectives (Liu 2004a):

- Minimize the investment cost of the units (i.e., surface area of exchanger, heater and/or cooler)
- Minimize the operating cost of utilities (e.g., steam, cooling, water, etc…)
- Minimize the number of units, N_{min} (i.e., exchanger, heater, and/or cooler)
- Achieving the most efficient, or nearly reversible exchanger of heat among hot and cold stream

The development of heat exchanger network starts with the collection of data. If the data is missing a quick heat balance or the use of simulation software like Aspen Plus© can be used to determine the missing data. Then the data should be sort as shown in the Table 4.3.

Table 4.3 Example of heat exchanger network problem

Stream	Capacity Flow Rate Cp	Input Temperature T_{in}	Output Temperature T_{out}	Heat to be added to cold stream, or heat to be removed from hot stream
Sc_1	Cp^c_1	$T^c_{in, 1}$	$T^c_{out, 1}$	$Q^c_1=Cp^c_1*(T^c_{out, 1}- T^c_{in, 1})$
Sc_2	Cp^c_2	$T^c_{in, 2}$	$T^c_{out, 2}$	$Q^c_2=Cp^c_2*(T^c_{out, 2}- T^c_{in, 2})$
…	…	…	…	…
Sc_i	Cp^c_i	$T^c_{in, i}$	$T^c_{out, i}$	$Q^c_i=Cp^c_i*(Tci^{out}-Tci^{in})$
Sh_1	Cp^h_1	$T^h_{in, 1}$	$T^h_{out, 1}$	$Q^h_1=Cp^h_1*(T^h_{in, 1}- T^h_{out, 1})$
Sh_2	Cp^h_2	$T^h_{in, 2}$	$T^h_{out, 2}$	$Q^h_2=Cp^h_2*(T^h_{in, 2}- T^h_{out, 2})$
…	…	…	…	…
Sh_j	Cp^h_j	$T^h_{in, j}$	$T^h_{out, j}$	$Q^h_j=Cp^h_j*(T^h_{in, j} - T^h_{out, j})$

A problem table for utility targeting composite stream analysis (Table 4.4) with a constant capacity flow rate and a global minimum approach temperature, ΔT_{min}:

Interval temperatures, T_b:

$$T_b = \left\langle T^C{}_i + \frac{1}{2}\Delta T_{\min} ; T^h{}_j - \frac{1}{2}\Delta T_{\min} \right\rangle \text{ with } i = 1...i \text{ and } j = 1...j \quad (4\text{-}6)$$

The cascade of net energy is derived from the composite interval diagram. One side it gives the boundary interval temperature including the pinch temperature, and the other side gives the heat required for each temperature interval (Fig. 4.3). At the beginning and the end, the minimum heat utilities are found.

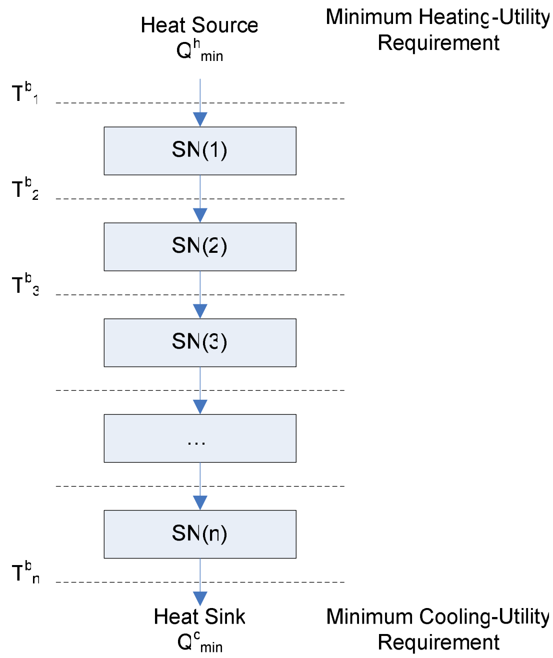

Figure 4.3 Cascade of the energy distribution of the HEN

The composite curve is plot from the composite interval diagram. By using the interval boundary temperature and the heat duty of hot and cold, it is also possible to plot the composite curve, which is shown in Fig. 4.4.

Table 4.4 The composite interval diagram

Interval Boundary Temperature T^b	Composite Hot Stream: Available heat	Composite Hot Stream: Cascaded heat	Hot Stream: **Sh_1** Cp^h_1	Hot Stream: **…**	Hot Stream: **Sh_j** Cp^h_i	Stream Temperature	Stream Temperature	Cold Stream: **Sc_1** Cp^c_1	Cold Stream: **…**	Cold Stream: **Sc_i** Cp^c_j	Composite Cold Stream: Available heat	Composite Cold Stream: Cascaded heat	Net Heat	Grand Composition Stream: Cascaded Heat	Grand Composition Stream: Adjusted Cascaded Heat
		0										0		0	$Q^c_{min}=0-Q_{min}$
T^b_1	Q^h_1					$T^h_{in,1}$	$T^c_{out,1}$				Q^c_1		$Q^h_{SN(1)}-Q^c_{SN(1)}$		
SN(1)		$Q^h_{SN(1)}$				$T^h_{in,2}$	$T^c_{out,2}$					$Q^c_{SN(1)}$		Q_1	Q_1-Q_{min}
T^b_2	Q^h_2						$T^c_{in,2}$				Q^c_2		$Q^h_{SN(2)}-Q^c_{SN(2)}$		
SN(2)		$Q^h_{SN(2)}$				$T^h_{out,2}$	…					$Q^c_{SN(2)}$		Q_2	Q_2-Q_{min}
T^b_3	Q^h_3						…				Q^c_3		$Q^h_{SN(3)}-Q^c_{SN(3)}$		
SN(3)		$Q^h_{SN(3)}$				$T^h_{out,1}$	…					$Q^c_{SN(3)}$		Q_{min}	0
…	…	…				…	….				…	…	…	…	…
…	…	…				…	$T^c_{in,1}$				…	…	…	…	…
T^b_n	Q^h_j					…	….				Q^c_i		$Q^h_{SN(n)}-Q^c_{SN(n)}$		
SN(n)		$Q^h_{SN(n)}$				$T^h_{out,j}$	$T^c_{in,i}$					$Q^c_{SN(n)}$		Q_n	$Q^h_{min}=Q_n-Q_{min}$

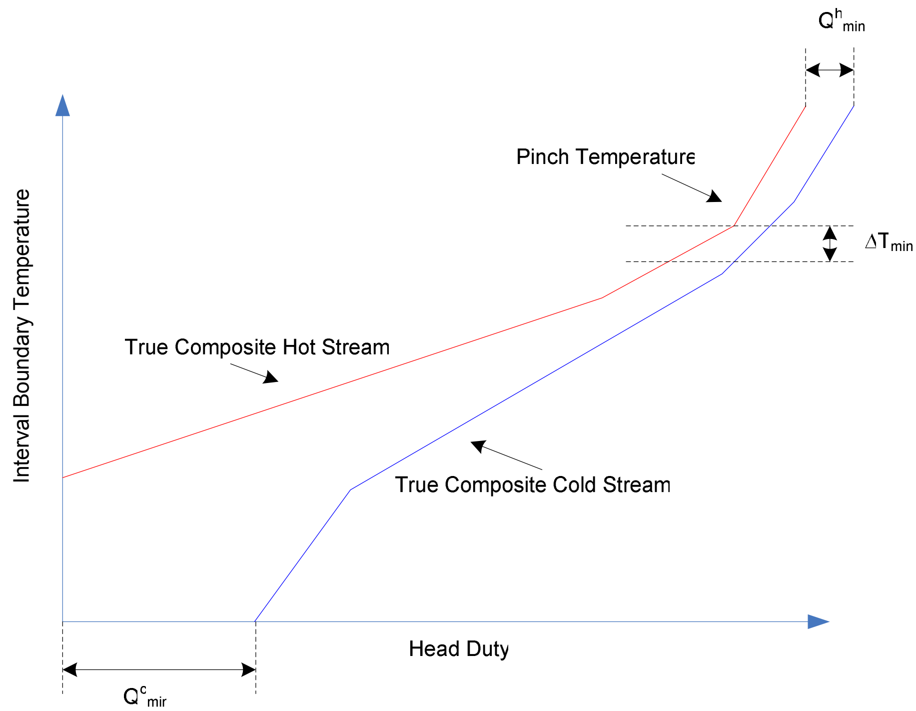

Figure 4.4 The composite curve

This graph provides valuable information as minimum heat source, heat sink and, the pinch temperature. This information will help to build the heat exchanger network. First of all, the minimum of number of units (exchanger, heater, and/or cooler) can be obtained by using the following equation:

$$N_{min} = N_h + N_c + N_{hu} + N_{cu} - 1 \qquad (4\text{-}7)$$

So far, three out of four targets have been determined, $Q^c{}_{min}$, $Q^h{}_{min}$, and N_{min}. The synthesis of an initial network is needed. One way to do it, is to use the composite curve and the enthalpy interval (Fig. 4.5). Every change on the true composite hot and cold stream is an enthalpy interval.

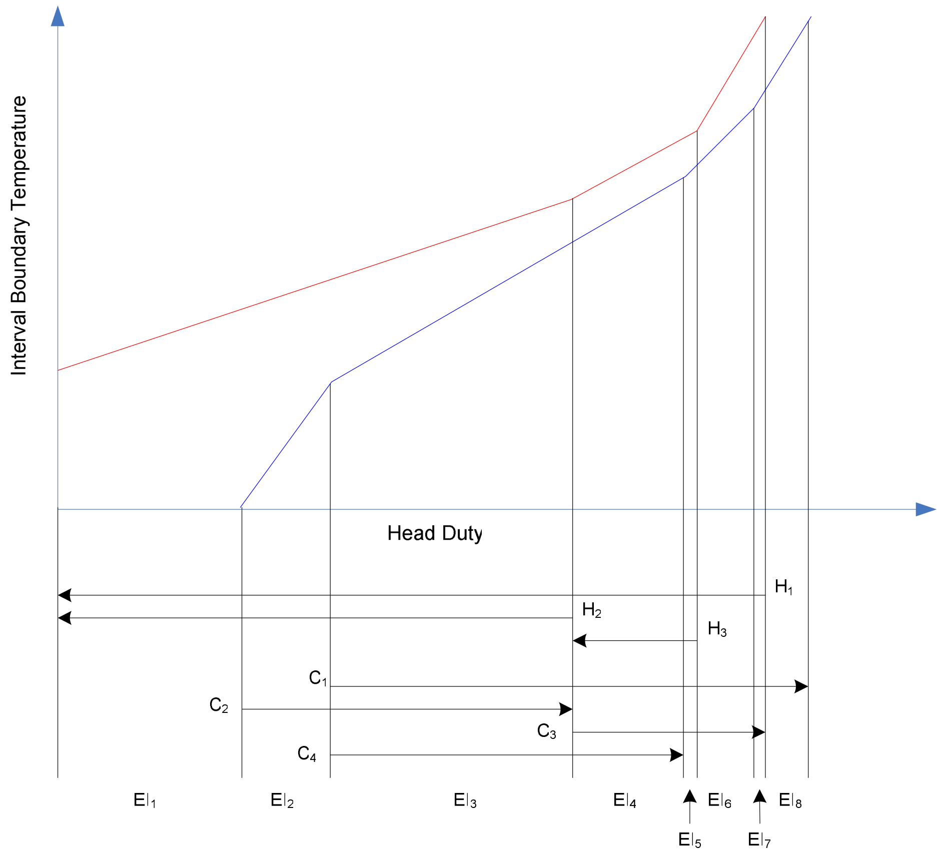

Figure 4.5 Description of the enthalpy interval method

Form this enthalpy interval method, it is possible to relate the hot interval to the cold interval. Then the initial network can be built (Fig. 4.6).

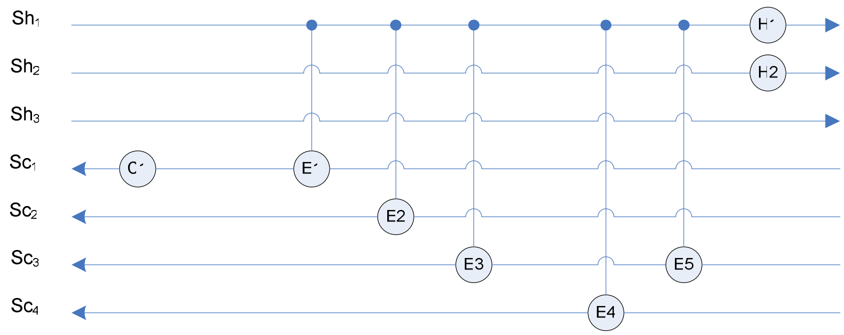

Figure 4.6 Initial network from the composite curve

An initial network usual has few violations like ΔT_{min}, Pinch overlap, or N_{min} not respected. These violations can be overcome by splitting the streams, “path” breaking or “loop” breaking. Once all issues have been addressed, the network is called improved network (Fig. 4.7).

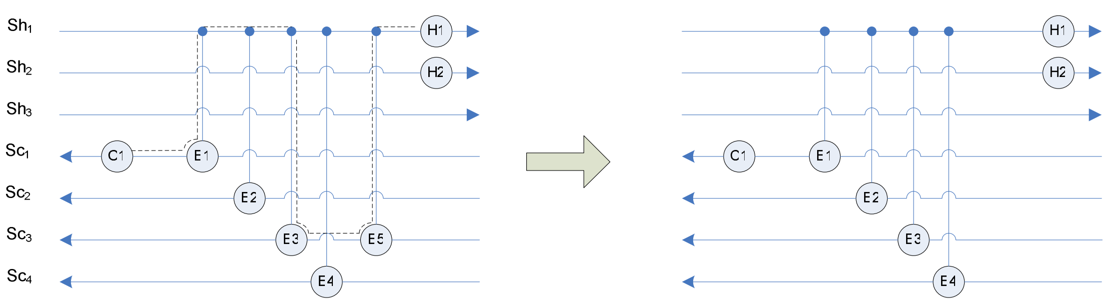

Figure 4.7 Improved network: Illustration of “path” breaking

The heat integration is only a tool that helps to achieve an optimal and feasible heat exchanger network. Because of a trade-off among minimum area, the minimum of the units and the pinch, sometimes the best heat exchanger network is not feasible.

4.1.3 Cost Analysis

For a new plant, the total cost investment is obtained from the equipment cost delivered by using several factors (Peters and Timmerhaus 2003). The equipment cost delivered is obtained from different resources as vendor, contractors, online (Vasudevan 2004), and handbook (Perry and Green 1997; Peters and Timmerhaus 2003). The operating costs are determined by market price of different chemicals used and the heat utilities cost (Liu 2004a). The income before tax is defined by subtracting the revenue to the operating cost. The depreciation is done over 20 years for a plant and 5 years for heat exchanger (Liu 2004b). Typically, a minimum of 15% for rate or return is needed to be an acceptable project although under current market conditions rates of return of 5-10% may be acceptable. This concludes the tools that are used in this work.

4.2 Description of the proposed flowchart

4.2.1 Define needs

The first step is to identify the project objectives. The needs of the company have to be well defined. Some key information are needed:

- Required capacity
- Feedstock constraints
- Total capital investment
- Company's short- and long-term objectives

- Time frame

4.2.2 Internal rearrangements

The internal rearrangement is the low cost option. This option is limited to the use of existing equipment by changing the process configuration through stream rerouting. Changes in operating conditions are also permitted. Furthermore, feedstock substitution is considered Mass and energy integration tools can be used to reconfigure the process and enhance material recovery, fresh-resource usage, waste discharge, and consumption of heating as well as cooling utilities. The yield can be modified to produce more out or the feed can be increased or changed to another one. The extra material produced will influence the rest of the process, either by increasing heat duty or residence time of the units. In the case of the new feedstocks, there are more constraints. First of all, the characterization the new feedstocks and the current plant technology are needed. For instance, if the plant is a biomass facility using corn, it would be easy to integrate another biomass feedstock. It would require a change in the procedure and modification in the operating conditions. In the other case when the plant is fossil-based technology and the company wants to integrate a biomass feedstock, there is a need to add new units and to change the operating conditions as well as stream routing. If the company desires to integrate the same kind of feedstock, internal rearrangement is possible. The next step is to simulate the possible candidate flowsheet using computer-aided simulation software. The proposed prototype can be improved by using mass and heat integration techniques. Once the optimal design is selected, a cost analysis can be performed. Based on the simulation, integration, and cost-benefit analysis, the optimal decisions can be made.

4.2.3 Internal modifications by adding new units

The internal modification is a medium cost alternative because it requires the addition of new units. The addition of a unit is particularly needed when the feedstock requires special pre-processing before it is rendered usable by the current plant. Once the new units are added, material rerouting and heat integration can be performed. The economics of the evolved process should be analyzed. It is possible to generate several alternatives due to the availability of different technologies. Once again, a profitability analysis is used to determine the optimal alternative.

4.2.4 External modifications by adding new lines

The external modification involves the addition of a new production line, possibly with a new technology. First, a review of all candidate processing technologies should be conducted. A preliminary techno-economic screening is preformed to narrow the selection to few candidates that warrant further consideration. In cases involving two candidates, they can be analyzed separately or merged with the current plant and analyzed. The higher rate of return on investment will determine the optimal option. The uniqueness of heat and mass integration proves over the last decade that they could give a different network configuration result depending on the initial inputs as minimum temperature approach, hot and cold stream distributions, etc. By varying the ratio between the two candidates and creating a hybrid plant (mixture of two candidates meeting required capacity), the initial inputs would change, (i.e. hot and cold streams distribution). Then the pinch analysis for each given mixture may give better or worst results than the candidates separately. Figure 4.8 represents the flowchart summarizing this design procedure.

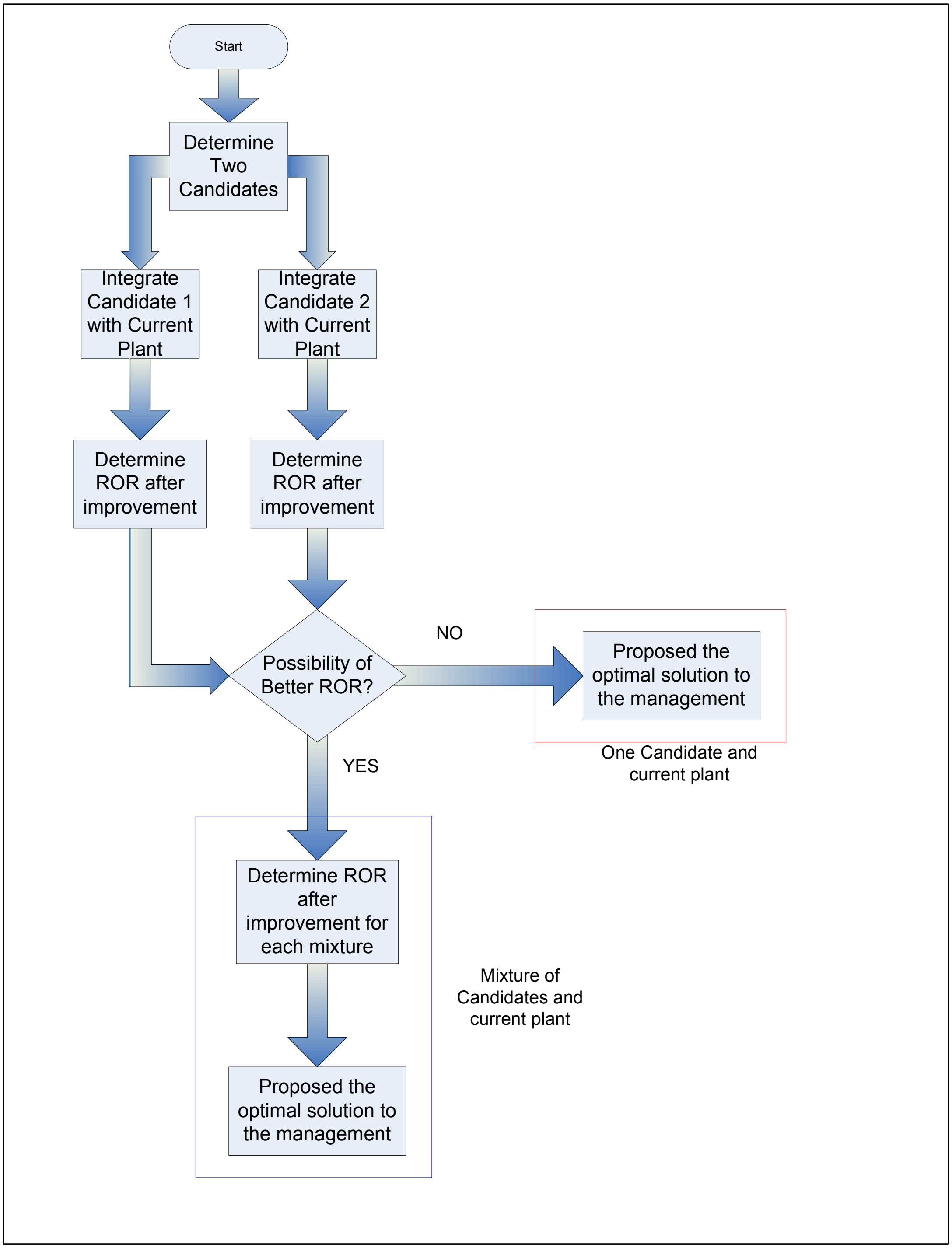

Figure 4.8 Proposed flowchart for external modification

Once the two candidates are integrated into the current plant by heat and mass integration, the rate of return can be calculated. The decision to go deeper into the analysis depends on the values of rate of return on investment. If there is great difference between these values, the highest value should be enough to decision on the optimal candidate. If the values are closed enough that an influence of configuration would make a significant improvement then the mixture of candidates should be considered and analyzed.

4.2.4.1 Graphical approach of ROR vs X

A graphical approach is used to determine the optimal solution in the case of comparing two new plants: one employing biomass and the other one employing fossil-based feedstock.. The variable X is introduced as the fraction of product produced by the biorefinery. The rest of the fraction represents the product fraction produced by the fossil-based plant. When the rates of return of each plant are comparable, additional investigation is needed. This is carried out by iteratively examining different values of X, simulating the processes, conducting heat and mass integration for each scenario, and conducting economic analysis. The result of this analysis is the identification of the optimal production capacity of each plant, the opportunities for heat and mass integration along with their implementation, and the performance of the combined process.

CHAPTER V

CASE STUDY: ETHANOL PRODUCTION

This research is aimed at approaching the issue of integrating a bio-refinery with a traditional processing facility. While the problem is general, attention will be given to the case of producing ethanol as an illustrating example. Consider a current fossil-based plant (ethylene feedstock) which produces ethanol. Because of the anticipated increase in ethanol demand, the plant is interested in increasing its production. To retrofit the plant, two options are considered: using the same ethylene-based technology while increasing capacity and/or adding a bio-refinery which uses fermentation of a biomass as feedstock. Both production lines should be integrated. Mass and energy integration will be used to reduce energy and material utilities and to identify shared equipment. As stated in the previous chapter, a systematic procedure is to be proposed to design the integrated facility and to assess the various options.

An existing 50 MMGPY ethanol plant is to be integrated with a new plant either ethylene or biomass. The current plant was uses ethylene as a feedstock. General information on this technology can be found in literature (e.g., Elkin et al., 1979). At present, the plant is looking to expand production to 80 MMGPY. The additional 30 MMGPY of ethanol may be of the fuel quality (through a biorefinery) or the chemical grade (from ethylene). The selling prices of the two product qualities are different with the chemical grade selling for higher prices than the fuel grade. Two alternatives are considered for increasing the capacity: another ethylene-based process and/or a biomass-to-ethanol process.

The objective of the case study is to retrofit the existing plant by one or both of the alternative technologies. The existing plant is assumed to be fully depreciated and paid for with the only remaining cost being the operating cost of the process. The new plant will incur a new total capital investment, operating cost, and sale of product. The main purpose of this case study is to perform an analysis on the two different technologies and make a choice based on the process economics.

5.1 Determination of technology and feedstock available for ethanol plant

The first step is to consider the candidate feedstocks. For the biorefinery, several feedstocks may be considered. These include

- Off-spec grain
- Molasses
- Sugar cane
- Barley
- Waste starches and sugars
- Ligno-cellulose

For illustration purpose, a lingo-cellulosic biomass is selected. It is assumed to be available at a cost of $20/dry ton. The same analysis can be repeated for other feedstocks. For the fossil-based facility, two routes may be considered for ethanol production. These include:

- Direct hydration of ethylene
- Carbon monoxide and hydrogen

For the fossil-based facility, the direct hydration of ethylene route is chosen. This is the same technology as the existing facility which currently produces 50 MMGPY of ethanol.

5.2 Current fossil refinery and candidate

Figure 5.1 is a schematic flowsheet of the existing ethanol plant that uses direct hydration of ethylene. They key aspects of the technology are described in literature (e.g., Elkin et al., 1979). The stream data of the ethylene plant for 30 and 50 MMGPY are tabulated in Appendix A. This plant has two sections, reaction and purification. Ethylene reacts with the water in the reactor to produce ethanol long with ether in the reactor area. Unreacted ethylene is separated using a scrubber and recycled back to the reactor. The product line which contains ethanol, water, and side product passes through a series of columns to obtain ethanol at the end. Water is recycled as well. Ethers separated and sold as a byproduct.

A computer-aided simulation tool, Aspen Plus©, is used to predict the key flows, compositions, equipment sizing which constitute the basis for evaluating the fixed cost and operating costs. To employ published fixed-cost data for plants of different capacities, the six-tenth factor rule is used, i.e.

$$\text{CostPlantB} = \text{CostPlantA} \times \left(\frac{\text{SizePlantB}}{\text{SizePlantA}} \right)^{0.6} \qquad (5\text{-}1)$$

For instance, Elkin et al. (1979) published cost data for an ethylene-to-ethanol process for a 30 MMGPY plant. Cost indices are used to update the cost.

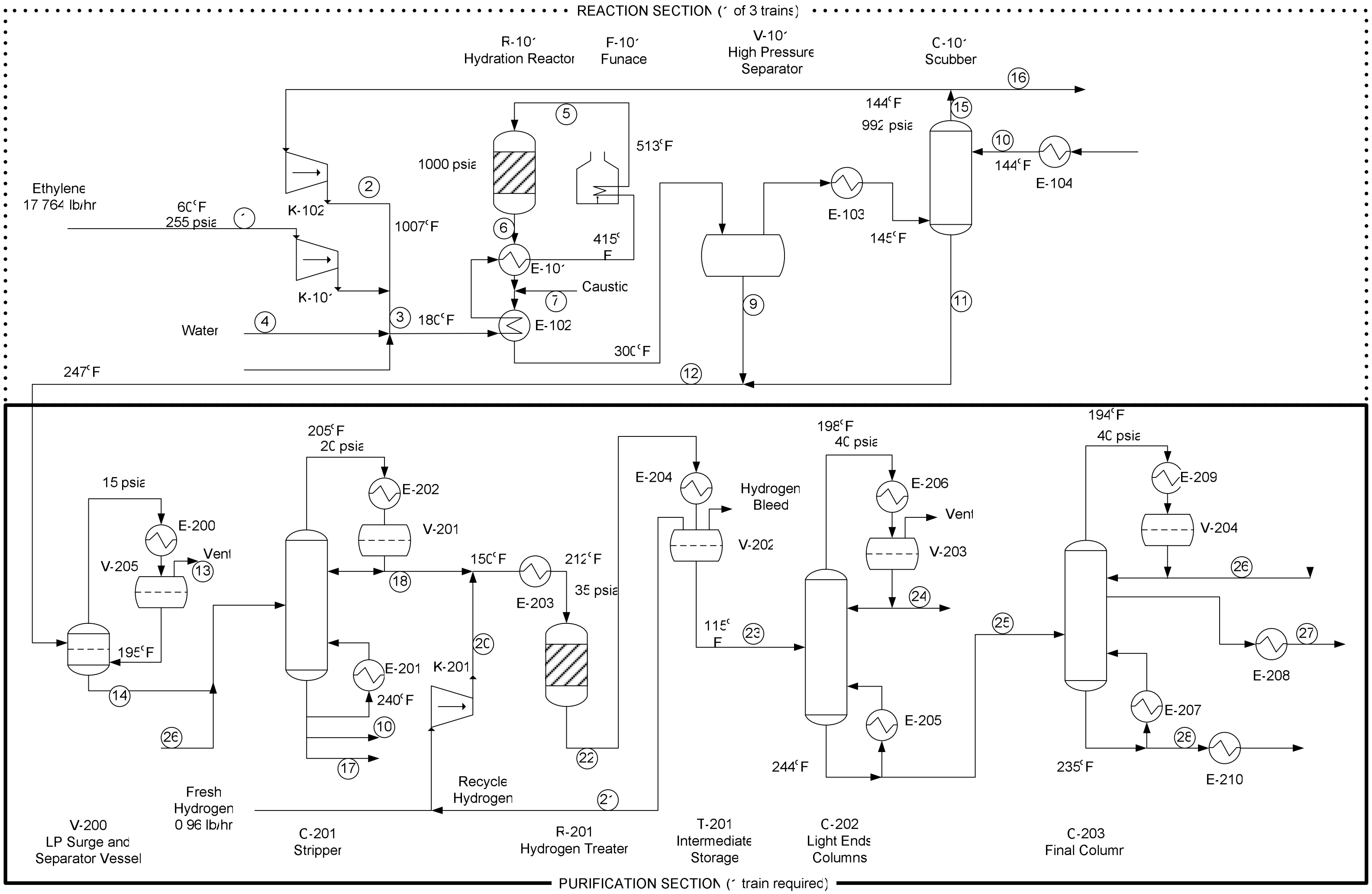

Figure 5.1 Flowsheet for ethylene plant

5.3 Bio-refinery candidate

The proposed biomass-to-ethanol on recent publications by NREL (Aden et al. 2002; Montague 2003). The ethanol plant uses corn stover as lignocellulose feedstock. Figure 5.2 is a schematic representation of the flowsheet. The lignocellulose feedstock was chosen because the flexibility that it provides. Other lignocellulose feedstock can be used with the same design. The adaptation of the different types of feedstock is only a matter of operating conditions.

The stream data of the corn stover plant for 30 and 69 MMGPY are tabulated in Appendix B. The flowsheet is broken down into three sections: pre-hydrolysis, saccharification and fermentation. Concerning the biomass chemistry involved in this section, the biomass materials can be broken down to six-carbon and five-carbon sugar. This sugar is transformed to ethanol by fermentation. The following are examples of the reactions:

Six-carbon sugar to ethanol

$$C_6H_{12}O_6 \rightarrow 2C_2H_5OH + 2CO_2 \qquad (5\text{-}2)$$

Five-carbon sugar to ethanol

$$3C_5H_{10}O_5 \rightarrow 5C_2H_5OH + 5CO_2 \qquad (5\text{-}3)$$

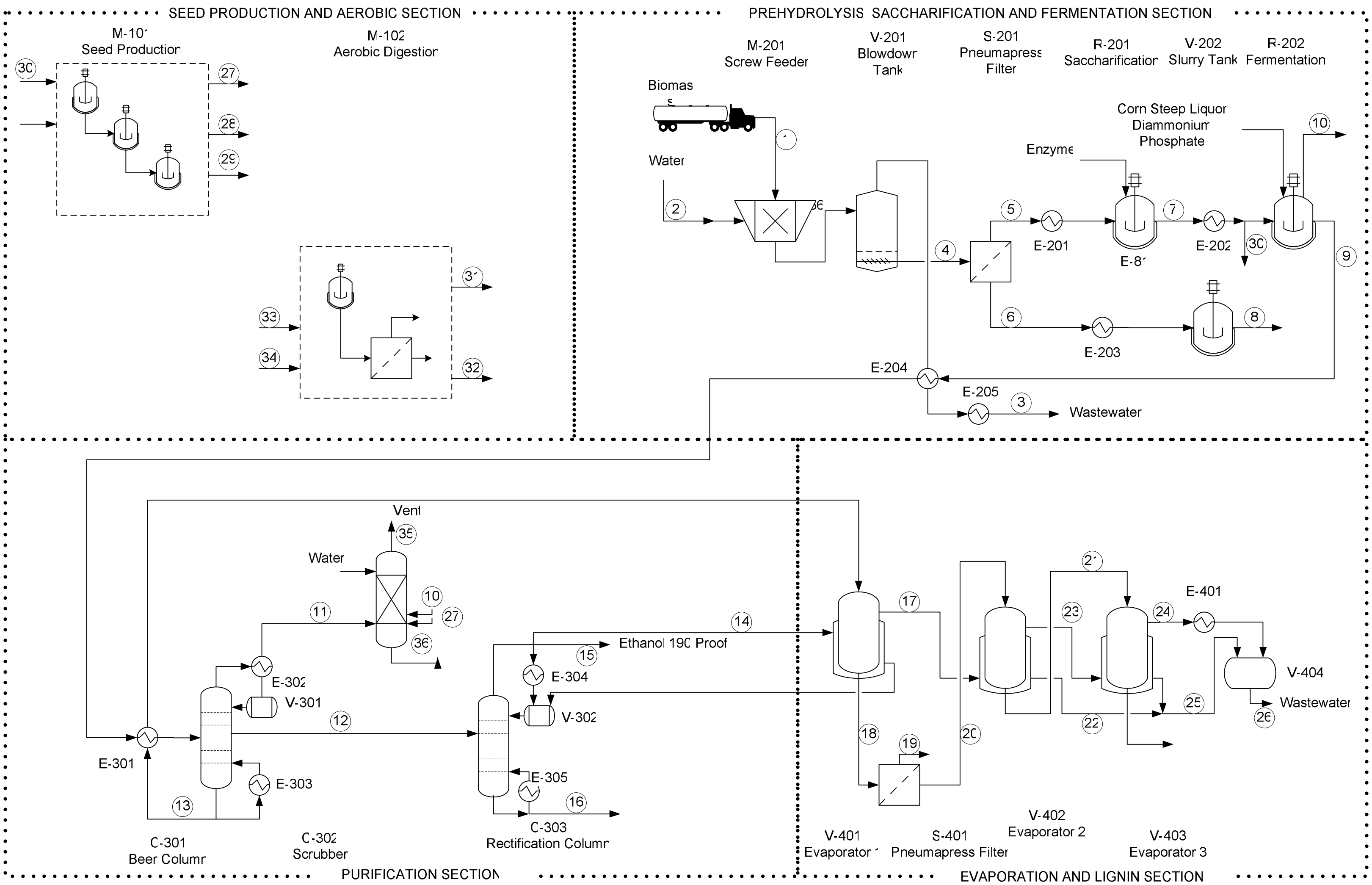

Figure 5.2 Flowsheet for corn stover plant

The actual reaction network may be quite complex involving many reactions that go through intermediate steps before yielding usable matter (e.g., Aden et al., 2002. The fermentation is a bio-chemical process. Enzymes, nutrients for the enzymes, and well water are needed in order to perform the fermentation. Typically, this procedure takes up to 24 hours to complete depending on the enzyme efficiency and the conditions. The seed production and aerobic section produce chemicals needed in the pre-hydrolysis, saccharification and fermentation section. The purification section separates ethanol/water from solids and then ethanol from water. The evaporation and lignin section separates solids from lignin and water.

The actual size of this plant described by Aden et al. (2002)is 69 MMGPY while this study needs a plant of 30 MMGPY. Again, Eq. 5.1 representing the six-tenth factor rule will be used to relate the fixed cost of the two capacities.

5.4 Using the proposed flowchart

To increase the capacity of ethanol production from 50 to 80 MMGPY, the proposed design approach described in Chapter IV will guide the solution in this section.

5.4.1 Internal rearrangements

Current ethanol plant: The existing ethylene plant may be improved using heat integration through the synthesis of a heat-exchange network. Integrating the biomass feedstock into the fossil-based facility is no practical because of the vastly-different units required for the fossil-based process versus the biomass-based process.

5.4.2 Internal modifications by adding new units

Ethylene: The integration of the ethylene plant option would require more than adding new units. The capacity expansion to 80 MMGPY would require an entire new plant.

Biomass: The biomass plant option requires the handling of solids and liquids. It requires external modifications.

5.4.3 External modifications by adding new lines

Ethylene feedstock: The expansion of the ethylene option requires three trains for the reactor section and one train for the purification section as presented in the SRI report (Elkin et al. 1979).

Biomass feedstock: Biomass would require an entire plant that is describe in the NREL report (Aden et al. 2002; Montague 2003).

5.5 Analysis of current ethanol plant

The existing plant produces 50 MMGPY of ethanol. This plant is paid for, which means there is no additional depreciation of the total capital investment. Only taxes and operating cost are subtracted from the sales to calculate the gross revenue. Throughout this case study, a 40% tax rate is assumed to apply to the gross revenue and the rest is the designated as the net revenue. The operating cost, revenue, and net revenue are summarized by Table 5.1.

Table 5.1 Economic data for 50 MMGPY current plant

Economic data for 50 MMGPY Existing Plant ($/yr)	
Operating cost	143,128,078
Sales	143,992,517
Net revenue	518,663

The operating cost can be divided into three subcategories, general operating cost, raw materials and utility operating cost. General operating cost is taken as a linear function of the plant size. General operating cost is regrouping salary, labs, and other costs. Raw material operating cost is also a linear cost and it includes the cost of all chemicals used in the plant for the production. Utility operating cost accounts for all heating and cooling utilities. It is taken as a linear function of the plant production rate. Table 5.2 represents the cost of each category of operating cost.

Table 5.2 Categories of the operating cost for the 50 MMGPYplant

Operating cost of each category ($/yr)	
General	281,598
Raw materials	96,991,440
Heat duty	45,855,040

The raw materials and the heat duty are the two most important operating costs of the current plant. The existing ethanol provides net revenue of $518,663 per year after tax. As stated previously, the current plant may be internally modified. These modifications are primarily in the form of mass and energy integration. Table 5.3 summarizes the results of the cost analysis of the proposed improvement.

Table 5.3 Cost analysis of the improvement for existing plant

Cost analysis of the improvement for 50 MMGPY ethanol plant	
Annual savings	$ 4,024,608 /yr
Investment	$ 2,304,833
Rate of return	184%

As shown, the current plant should accept the proposed modification. The rate of return is very high. This rate of return was calculated over a period of 5 years for depreciating the capital investment. Table 5.4 shows the improvement of the revenue after modifications

Table 5.4 Comparison of net revenue before and after optimization

Comparison of old revenue and new revenue for 50 MMGPY ethanol plant	
Old	$ 518,663 /yr
New	$ 4,543,271 /yr

This modification will increase the net revenue by a factor of 8.7. The improvement will help in the future analysis. The details of the cost analysis are given by Appendix A.

5.6 New ethylene feedstock candidate

A new ethylene-based plant is also considered for increasing the ethanol capacity by 30 MMGPY. The basis for design is the work of Elkin et al. (1979). However, the design was improved by using heat integration.

Figure 5.3 shows the key sections of the ethanol plant with ethylene feedstock. Every section has a detail about their functions and the heat duty requirement to meet their functions. The first box sums up the heat duty requirement. The second box sums up the minimum heat utilities required as reported by Elkin et al. (1979). The last box sums up the minimum heat utilities found during the heat integration performed using thermal pinch analysis. Table 5.5 compares the savings resulting from process integration.

Table 5.5 Cost versus savings of heat integration for the 30 MMGPY ethanol-from-ethylene plant

HEN savings for 30 MMGPY ethylene feedstock	
Fixed cost	$ 76,022
Annual savings	$ 926,856/yr

The heat-integration savings is about a million dollars per year compared to the old design. Operating cost, taxes, depreciation of the total capital investment is subtracted from the revenue to obtain the net revenue. Table 5.6 shows the calculation of the income.

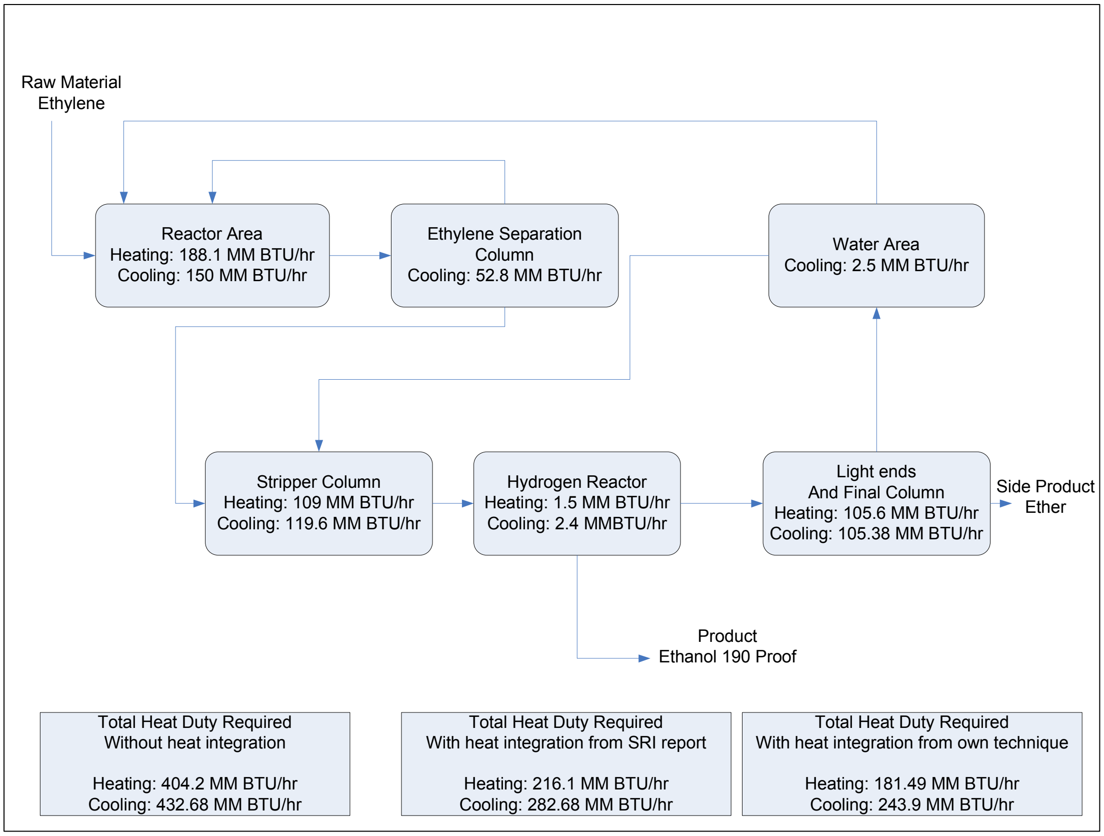

Figure 5.3 Heat duty required for each design of ethylene feedstock

Table 5.6 Economic data of 30 MMGPY new ethylene-based plant

Economic data for 30 MMGPY new ethylene plant	
Total capital investment	$ 66,265,629
Operating cost	$ 81,852,239/yr
Sales	$ 86,395,510/yr
Net revenue	$ 2,725,963 /yr
Rate of return	2.56%

The next ethylene plant will generate a net revenue of $ 2,725,963 per year. The operating cost is the same as describe before. Table 5.7 sums up the different categories.

Table 5.7 Operating cost of each category of 30 MMGPY new ethylene-based plant

Operating cost of each categories ($/yr)	
General	168,959
Raw materials	58,194,864
Heating and cooling utilities	23,488,416

The details of the cost analysis are reported in Appendix A2.

5.7 New biomass feedstock candidate

The ethanol plant form biomass is designed based on data by Aden et al. (2002). This design is for a production of 69 MMGPY of ethanol. The biomass feedstock is corn stover. Equation (5.1) is used to estimate the total capital investment from 69 MMGPY to 30 MMGPY of ethanol. The price of corn stover biomass is taken as $20 per dry ton. The heat utility is also improved from the previous design by using thermal pinch analysis for heat integration. The distribution of the heat duty required for each section is shown in Fig. 5.4.

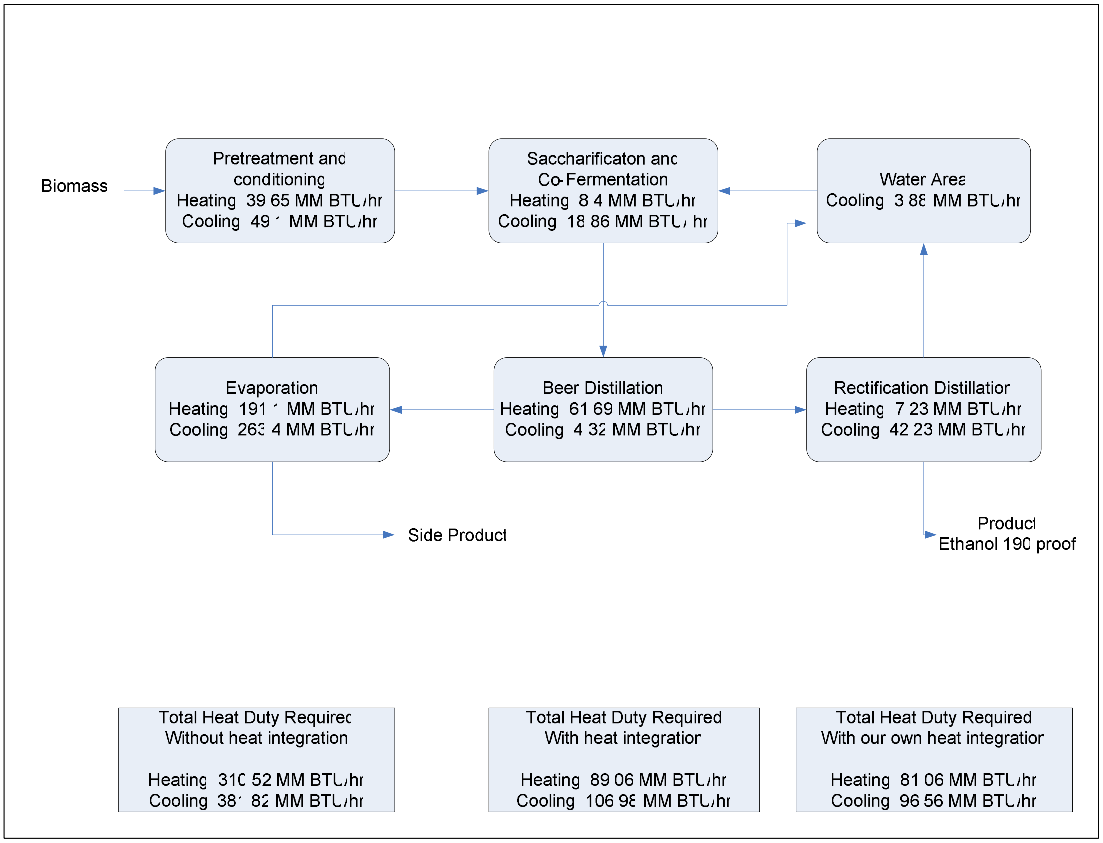

Figure 5.4 Heat duty required for biomass plant 30 MMGPY

As used before, the same boxes are used for this design. Table 5.8 shows the details of the old design and the improvement of the new design. The increment of this change is also calculated.

Table 5.8 Savings generated by HEN optimization

HEN savings for 30 MMGPY biomass feedstock	
Fixed cost	$ 228,136
Annual savings	$ 4,024,608/yr

Additionally, heat integration is conducted within the bio-refinery to reduce the operating cost. The total capital investment, operating cost, revenue and rate of return are summarized in the Table 5.9.

Table 5.9 Economics for the 30 MMGPY ethanol-from-biomass plant

Revenue for 30 MMGPY new biomass plant	
Total cost investment	$ 140,130,093
Operating cost	$ 16,485,454 /yr
Sales	$ 38,758,621 /yr
Net revenue	$ 13,363,900 /yr
Rate of return	9.98%

As can be seen, the bio-refinery is more expensive to build than ethylene-based plant but it offers a higher rate of return. The rate of return of the biomass plant compared to the ethylene plant explains why the majority of ethanol (for fuel purposes) plants in the world use biomass as feedstock. Ethylene is still used in the world (5% synthetic ethanol and 95% fermentation ethanol). However, it is primarily used for specialty applications that require certain specifications (e.g., cosmetics, food additives, etc.). Table 5.10 shows the detail of the operating cost for the biomass plant.

Table 5.10 Operating cost for the 30 MMGPY ethanol-from-biomass plant

Operating cost of each category ($/yr)	
General	5,923,913
Raw materials	12,608,696
Heating and cooling utilities	4,257,193

The details of calculation of the cost analysis are reported in Appendix B2.

5.8 Integration of two plants

In this section, the analysis of each plant separately is first reviewed. Then, each candidate is linked to the current ethanol plant by heat and mass integration. The quality of ethanol is different depending on the technology used. In the case of ethylene-based refinery, the produced ethanol is called chemical ethanol. It is primarily used for specialty applications (e.g., cosmetic sprays, food additives, etc.) and is sold at a price of $2.80 per gallon. In the case of bio-refinery, the produced ethanol referred to as fuel ethanol. It is used primarily for energy applications and is sold at a price of $1.21 per gallon.

5.8.1 In the case of separate plants

In this part of the analysis, the operating cost and the revenue of the existing plant are incorporated in the rate of return of each new plant. Table 5.11 shows these results.

Table 5.11 Rate of return for the new plant candidates

Rate of return for new plants	
Biomass	9.98%
Ethylene	2.56%

The rate of return of ethylene-based and biomass-based facilities is approximately 3% and 10%. The ethylene-based rate of return is lower than the typical minimum acceptable rate while the 10% for the bio-refinery is acceptable under current market conditions. The biomass looks to be the most attractive solution so far. The raw material is the most important operating cost for the ethylene plant. Ethylene becomes expensive with the increase in crude oil prices. From this result, the biomass is selected to be the feedstock to the new plant which produces 30 MMGPY in the case where the plants are separated. The next step considers the new plant along with the existing plant.

5.8.2 Integrating the two plants (hybrid plants)

In this part, the new plant option is added to the current plant. First, a 30 MMGPY ethylene-based plant is added to the existing 50 MMGPY ethylene-based plant. Next, a 30 MMGPY biomass-based plant is added to the existing 50 MMGPY ethylene-based plant.

In the case of adding an ethylene-based plant, except for the mass and heat integration, the configuration of ethanol plant with ethylene feedstock should be not change much. The new line is merged with the existing plant. The main functions and different sections are represented in Fig. 5.5.

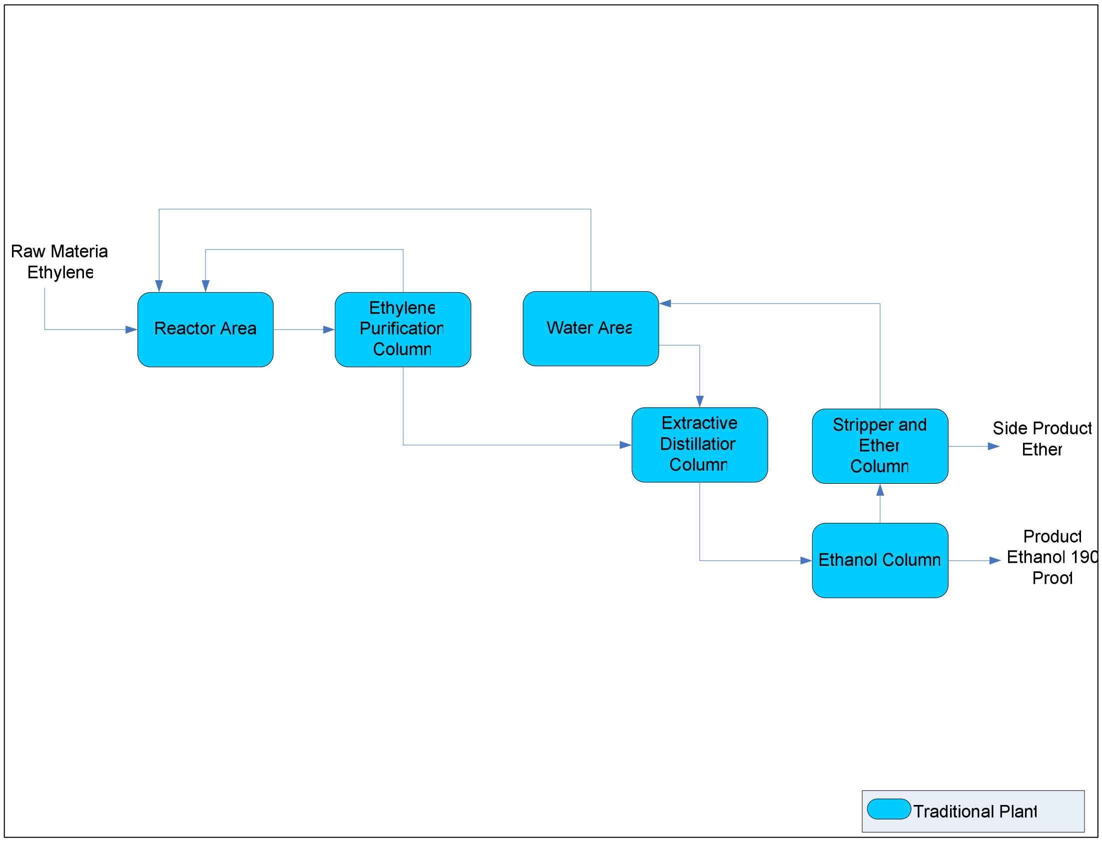

Figure 5.5 Configuration of expanded ethanol plant with ethylene feedstock

The configuration of the current plant is upgrade to 80 MMGPY. A summary of process economics is shown in Table 5.12

Table 5.12 Economics for 80 MMGPY hybrid plant with fossil/fossil feedstock

Economic data for 80 MMGPY ethylene-ethylene plant	
Total capital investment	$ 66,265,629
Operating cost	$ 218,272,637/yr
Sales	$ 230,388,028/yr
Net revenue	$ 7,269,234/yr
Rate of return	12.23%

In the case of adding a biomass-based plant, some of the sections of the current ethylene place replace some of the sections of biomass. In Fig. 5.6, the hybrid plant is presented.

In this case, the water and the purification sections are used by both the ethylene and the biomass plants. Because of the different qualities and markets of the biobased ethanol and the ethylene-based ethanol, the two different ethanol products have to be separated and stored in different units. There are two byproducts: syrup and ether, which are sold on the market. The current plant would produce 50 MMGPY and the biomass plant would produce 30 MMGPY. The return on investment for the bio-fossil integrated plant is about 14%. These results are presented in Table 5.13

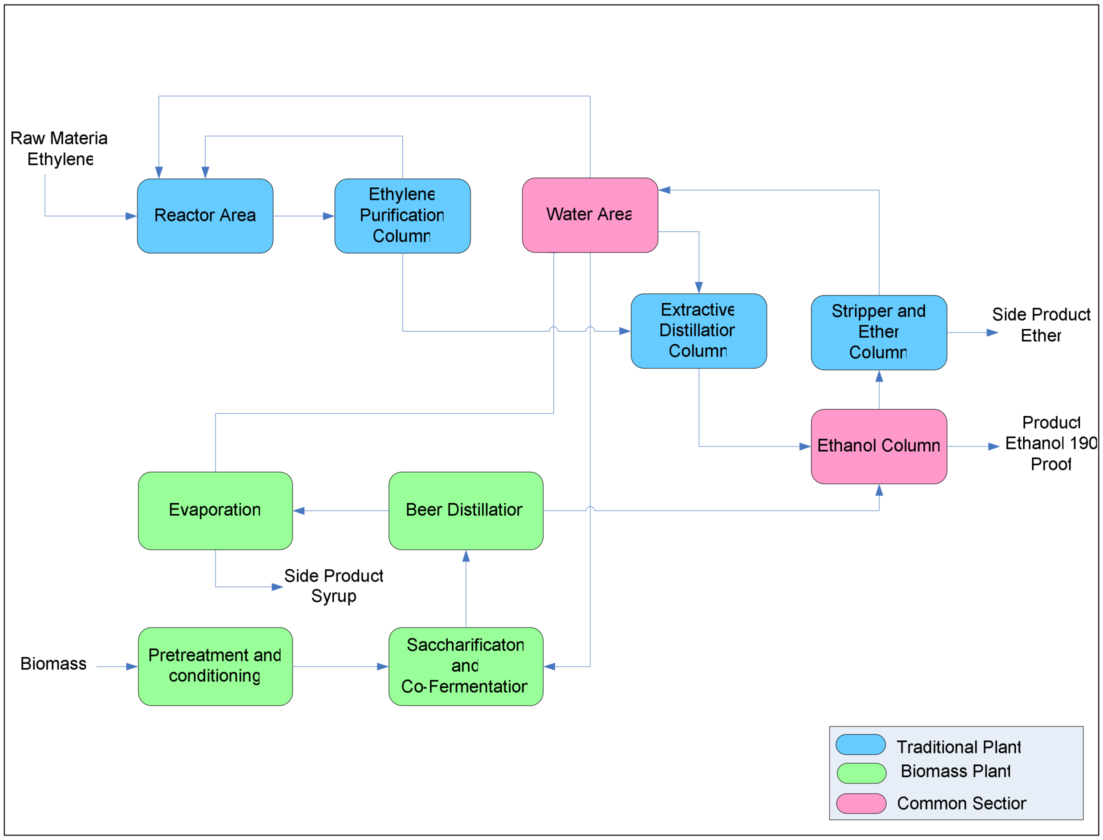

Figure 5.6 Configuration of ethanol hybrid plant with bio/fossil feedstock

Table 5.13 Economics for 80 MMGPY hybrid plant with bio/fossil feedstock

Economics for the 80 MMGPY bio-ethylene hybrid plant	
Total cost investment	\$ 140,130,093
Operating cost	\$ 152,905,852 /yr
Sales	\$ 182,751,138 /yr
Net revenue	\$ 17,907,171 /yr
Rate of return	13.91%

Table 5.14 summarizes the rates of return for the two alternative plants. As can be seen, the bio-refinery with the exiting ethylene-based facility yield a slightly higher rate of return than integrating a new ethylene-based plant with the existing ethylene-based facility. Additionally, there are environmental benefits that accrue as a result of using the biorefinery. Particularly, the net emission of green-house gases (primarily CO_2) is reduced from a life cycle perspective when the bio-refinery is used. Therefore, the scenario of the hybrid biorefinery-ethylene based facility is further considered by examining the effects of CO_2 benefits and by conducting heat and mass integration.

Table 5.14 Rate of return from the hybrid plant candidates

Rate of return for new plants	
Biomass	13.91%
Ethylene	12.23%

5.8.3 Heat and mass integration for the bio-refinery-ethylene hybrid plant

The economics of the 30 MMGPY new biorefinery and the existing ethylene-based plant can be improved by pursuing mass and heat integration between the two plants. The graphical pinch analysis targeting for material recycle/reuse network can be performed on water. Table 5.15 shows the distribution of water needed for a variety of columns present in the facility.

Table 5.15 Water dats of sinks and sources

Sink	Flow rate lb/hr	Mass Fraction	Load lb/hr
R-101	178,500	0.002	357
C-101	51,300	0.01	513
C-302	33,000	0.1	3,300

Source	Flow rate lb/hr	Mass Fraction	Load lb/hr
C-201	202,800	0.0017	345
C-203	6,450	0.018	116
C-303	34,000	0.004	136

The terms R-101, C-101, C-201, C-203 refer to the reactor and the columns in the current ethylene plant, and C-302 and C-303 are from the biomass feedstock. Figure 5.7 is the material recovery pinch analysis for the process.

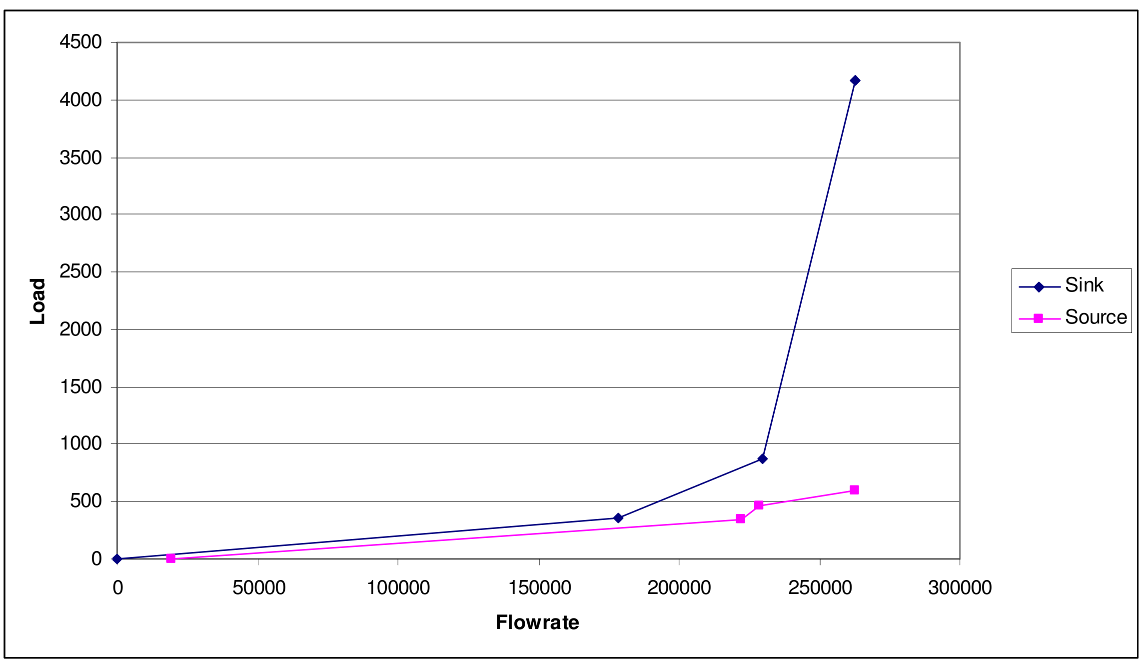

Figure 5.7 Material recovery pinch analysis for the process

The minimum fresh water required is 19,550 lb/hr instead of 140,100 lb/hr. This saving represents $ 50,631 per year.

The heat exchanger network can be synthesized in a similar way. The hybrid plant is composed of a total of 28 hot and cold streams. Above the pinch the minimum units should be 18 units and below the pinch should be 21 units. The data for the hot and cold streams are summarized by Table 5.16.

By using the techniques described in the Chapter IV, an optimum heat-exchange network is synthesized. The pinch point is at 193^{o}F for the hot streams and 180^{o}F for the cold streams. The minimum target heating utility is 505.35 MM Btu/hr and the minimum target cooling utility is 403.53 MM Btu/hr. Figure 5.9 shows the design of the heat exchanger network. The minimum heating and cooling utility are met while respecting the pinch. The minimum number of heat exchangers is used to reach the utility targets. The details of the pinch diagram analysis are shown in Appendix C.

Table 5.16 Data for the hot and cold streams

Hot stream	T_{Supply} ^oF	T_{Target} ^oF	Heat duty MM $Btu.hr^{-1}$	Specific heat MM $Btu.hr^{-1}.^oF^{-1}$
Sh1	541	300	125.00	0.52
Sh2	300	145	88.00	0.57
Sh3	195	80	21.00	0.18
Sh4	193	170	178.33	7.75
Sh5	212	115	4.00	0.04
Sh6	217	190	5.67	0.21
Sh7	182	100	4.27	0.05
Sh8	183	182	266.7	266.7
Sh9	218	135	1.01	0.01
Sh10	235	140	0.63	0.01
Sh11	214	86	45.09	0.35
Sh12	131	122	4.21	0.47
Sh13	149	106	8.44	0.20
Sh14	138	95	3.87	0.09
Sh15	149	145	91.85	22.96

Cold stream	T_{Supply} ^oF	T_{Target} ^oF	Heat duty MM $Btu.hr^{-1}$	Specific heat MM $Btu.hr^{-1}.^oF^{-1}$
Sc1	180	415	125.00	0.53
Sc2	144	226	4.17	0.05
Sc3	217	226	181.67	20.19
Sc4	166	212	2.50	0.05
Sc5	224	225	11.00	11.00
Sc6	200	218	264.00	14.67
Sc7	235	253	58.26	3.24
Sc8	106	203	38.83	0.40
Sc9	124	149	14.53	0.58
Sc10	203	212	3.76	0.42
Sc11	171	189	58.85	3.27
Sc12	142	160	95.76	5.32
Sc13	130	147	91.54	5.38

The benefit of integrating the biorefinery with the existing plant can be calculated for the heat exchanger network. Assuming, a $6/MMBtu of heating utility, the difference between integrated refineries and the un-integrated refineries $6.5 MM/yr of savings,

Figure 5.8, illustrates the savings as a function of the unit cost of heating utility. The cooling utility is assumed to cost 1.16 times the cost of the heating utility.

Appendix C shows the details of screening heat integration over various mixing ratios of the two plants using an increment of 10%. During this analysis, it is found that the optimal heat integration is around a 20% of ethylene feedstock option mixed with 80% of biomass feedstock option.

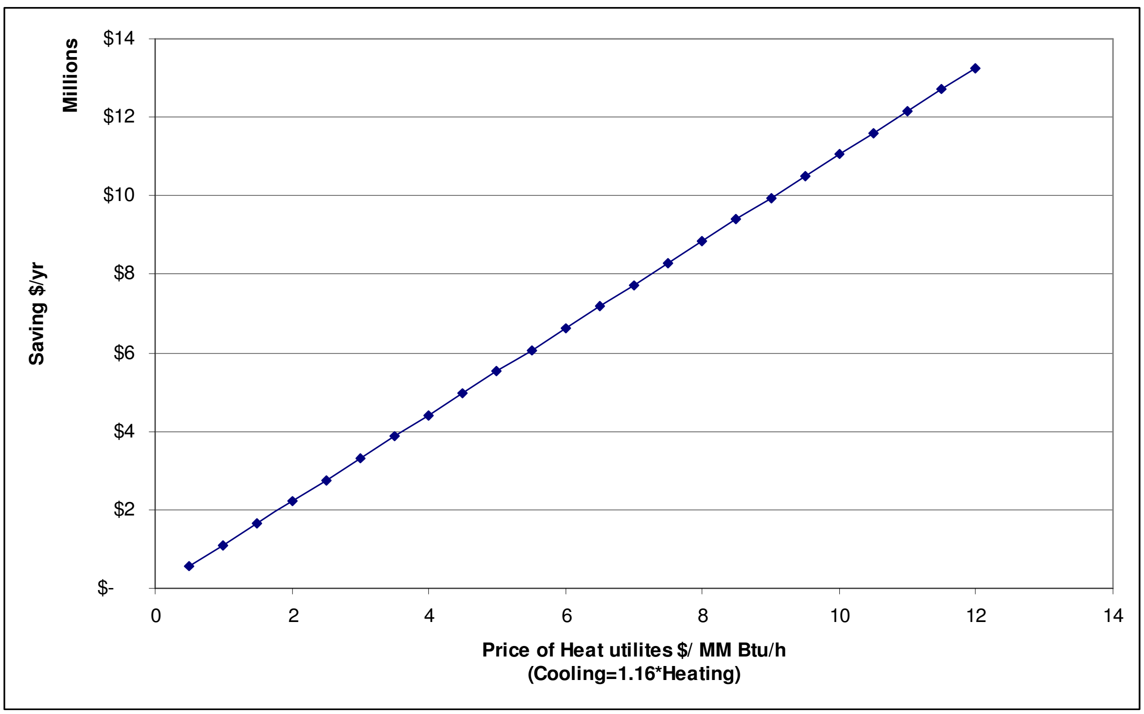

Figure 5.8 Saving of the integrated hybrid plant compare to the no integrated hybrid plant

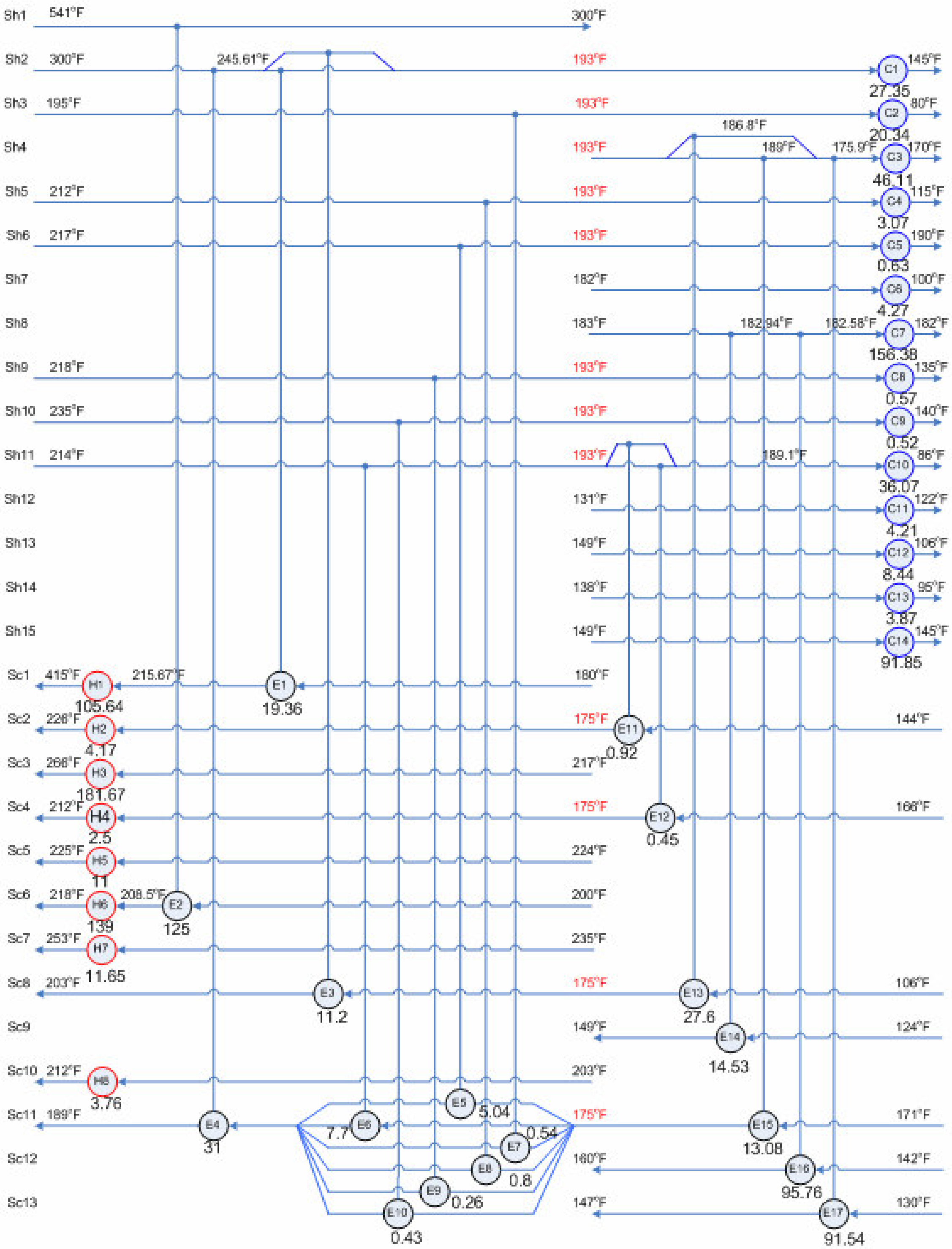

Figure 5.9 Heat exchanger network

5.8.4 Possible CO_2 regulations

The use of fossil-based fuel increases the greenhouse gas (GHG) emissions. With growing public concerns and international pressures over GHG emissions, it is possible that the US may start imposing GHG-emission penalties or providing credit to facilities discharging GHGs as a function of how the performance compares to certain benchmarks. Table 5.15 represents the effects of incorporating GHG penaltiy/credit on the cost of ethanol. Assessing the taxation of greenhouse effects is based on producing ethanol from corn, which has a net positive release of carbon dioxide of 6 kg/MM Btu/hr. We assumed for this case that the bio-refinery feedstock is herbaceous and it has a net negative release of carbon dioxide of 5.5 kg/MM Btu/hr. For the ethylene-based plannt, the net positive release of carbon dioxide is 42 kg/MM Btu/hr. We also assumed that the carbon dioxide is charged at $2/metric ton.

The margin for bio/fossil refinery has a greater margin than the fossil/fossil refinery. Clearly, as the GHG credits increase, the bio-refinery will be making additional profit.

Table 5.17 Cost of a gallon of ethanol with GHG penalty

	Fossil plant 80	Hybrid 50/30	Biomass plant 80
Operation cost	$ 218,272,637	$ 152,905,852	$ 36,741,573
Annual fixed cost	$ 3,313,281	$ 7,006,505	$ 10,833,233
GHG penalty	$ 857,143	$ 433,036	$ (273,810)
Total annual cost	$ 222,443,062	$ 160,345,392	$ 47,300,997
Ethanol cost $/gal	2.78	2.00	0.59
Margin $/gal	0.03	0.21	0.62

5.9 Sensitivity analysis

In this section, different variables were modified in order to study the behavior of the rate of return of each option. The understanding of the sensitivity of the rate of return helps in making the optimal decision under given conditions.

5.9.1 Variation of ethanol price

In this study, there are two different ethanol products (chemical and fuel). Their specifications and market use are different. In Fig. 5.10, the price of fuel ethanol was changed from 1.21 to 2.81 $/gal for fuel and chemical ethanol, respectively. Each company has its own minimum acceptable rate of return. Typically, a minimum of 5-10% rate of return is needed to further consider the project. In some cases, a minimum of 15% rate of return is taken as the minimum attractive rate. All the examined prices of biomass feedstock (from $20 to 40/dry ton) provide a return on investment higher than 15% when the price of fuel ethanol is at least $1.4/gal.

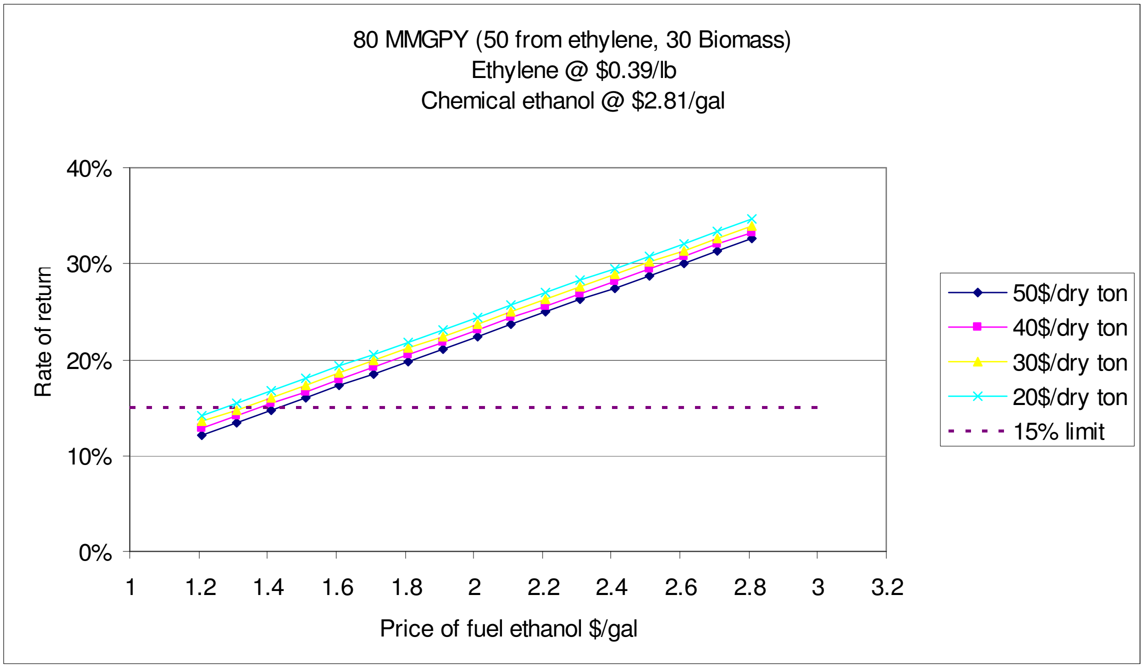

Figure 5.10 Variation fuel ethanol with ethylene feedstock constant

In the case of fossil/fossil refinery, the price of chemical ethanol was changed from 1.21 to 3 $/gal in order to make the comparison with Fig. 5.10. In Fig. 5.11, the 15% rate of return limit is drawn to illustrate the feasible combinations.

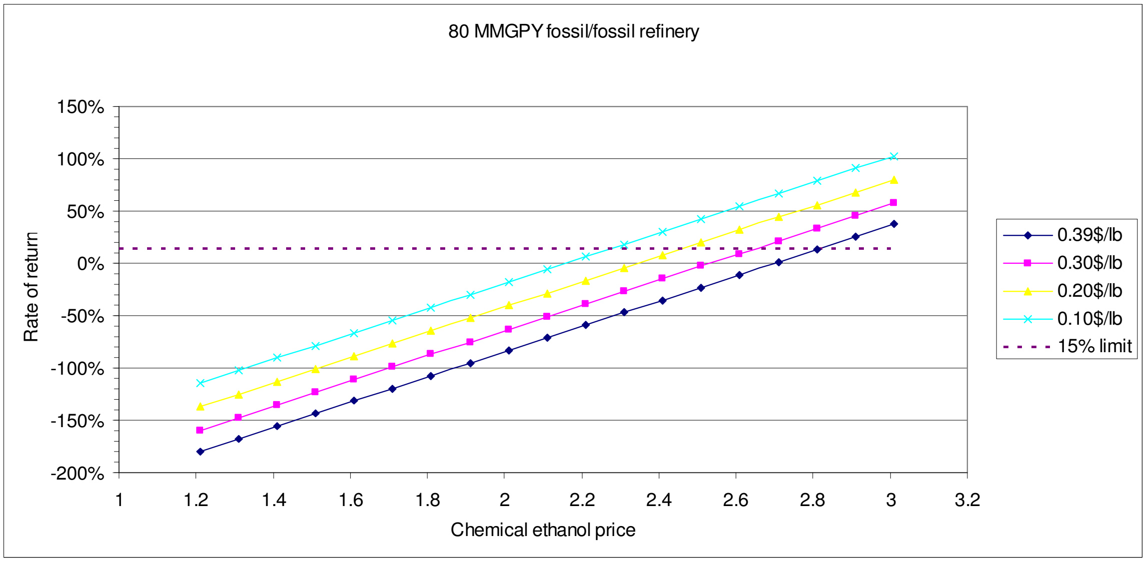

Figure 5.11 Variation of chemical ethanol price ($/gal) with biomass feedstock price

5.9.2 Variation of feedstock and utility prices

A sensitivity analysis is performed by changing the operating cost variables such as market price of raw materials, cooling and heating prices. The variables used for the sensitivity analysis vary from a location to another and a time to another. Table 5.18 summarizes the variables used in the analysis.

Table 5.18 Variable used in the sensitivity analysis

Variable	Ethylene	Biomass	Steam	Cooling	Process water	Chemical ethanol	Fuel ethanol
	0.39$/lb	$40/Dry Ton	7$/MM Btu/hr	6$/MM Btu/hr	5E-5$/lb	2.81$/Gal	1.21$/Gal
U.S.	100%	100%	100%	100%	100%	100%	100%
Utilities	100%	100%	150%	10%	100%	100%	100%
Middle East	60%	500%	80%	50%	50%	100%	100%

Table 5.18 presents all the variables that are changed during this analysis. Three variations are aimed at studying the following cases: typical operation in the U.S, high steam price, and operation in the Middle East. For each case, the sales, operating cost and rate of return are calculated. The analysis is also extended to the mixture by increment of 0.1. Figure 5.12 shows the rate of return as a function of the fraction of ethanol produced by the biorefinery out of the 30 MMGPY additional capacity (the rest of the fraction represents ethanol from ethylene).mixture of technologies for each case.

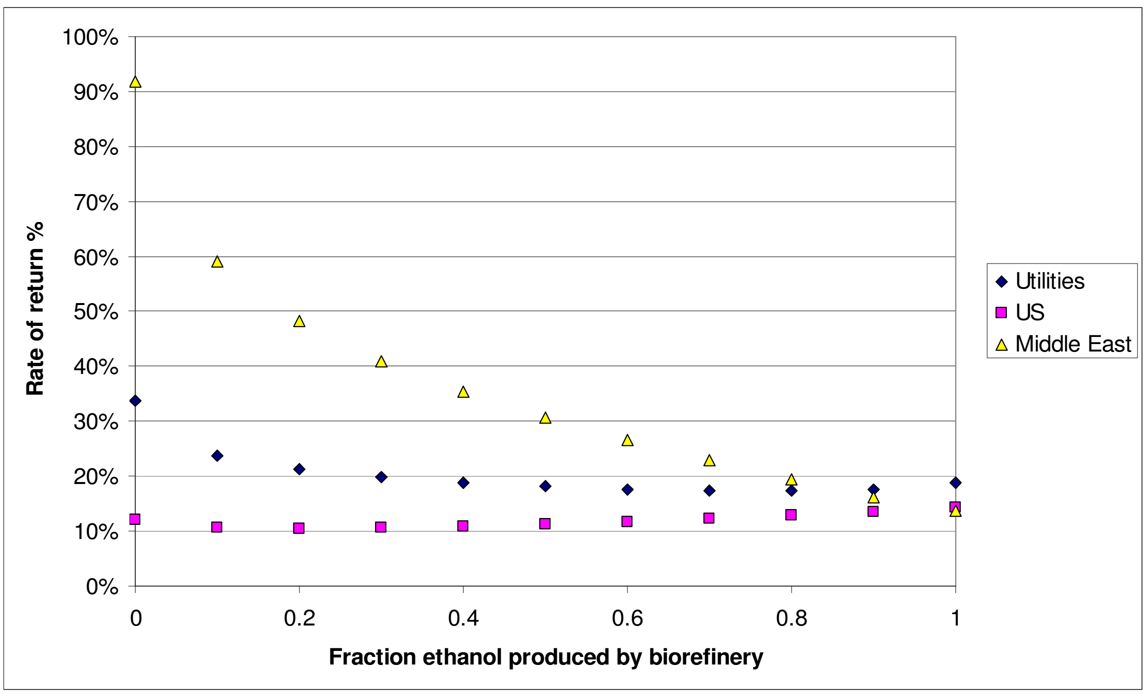

Figure 5.12 Rate of return as a function of the fraction of ethanol produced by biomass

This graph shows the three different cases. For the U.S. case, bio refinery is the optimal option. In the case of utilities, the fossil refinery is the optimal option. In the case of the Middle East, the ethylene option is more attractive and the revenue is high. In this case, it would be advisable to choose the ethylene option. More details are given by Appendix D.

For the calculation of rate of return at each step, the revenue, operating cost, and total capital investment were obtained with a linear interpolation. Heat integration and the grand composite curve were evaluated for each mixture for an increment of 0.1. For each step, the heat exchanger technique gave a better heat integration. The comparison of the linear interpolation and the calculated method and the savings of operating cost and total capital investment are represented in Fig. 5.13.

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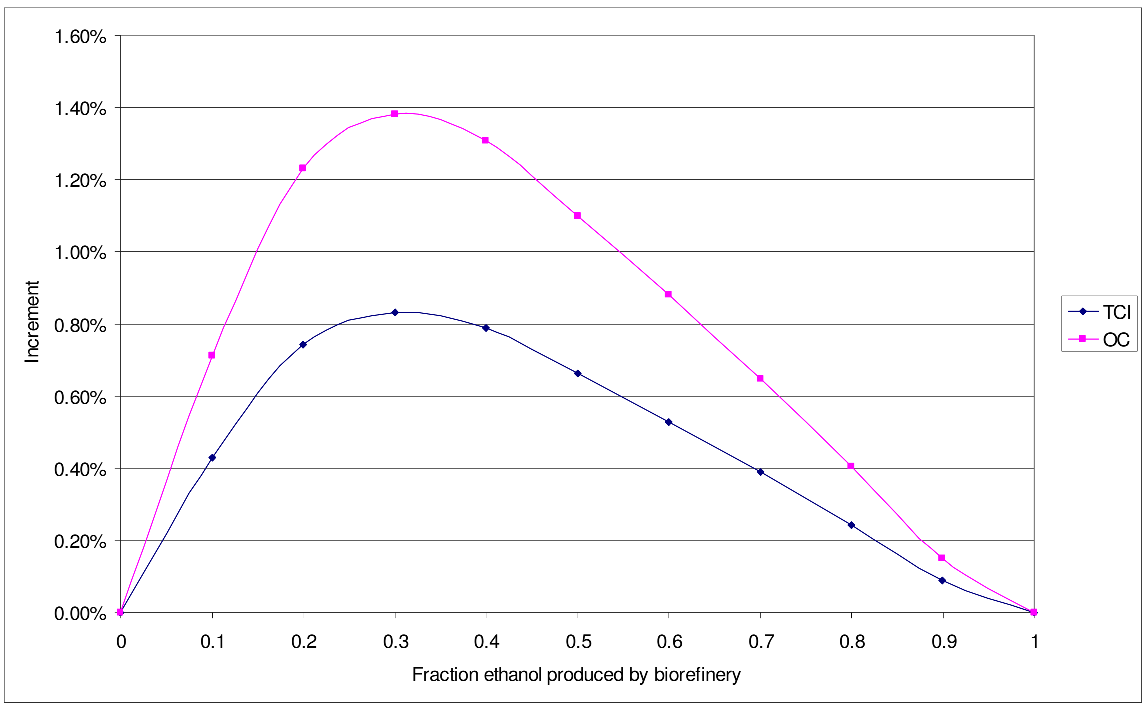

Figure 5.13 Comparison of increment of rate of return between HEN calculated and linear

If heat integration was the only concern, the optimal hybrid plant would be 70% of the ethylene feedstock and 30% of biomass feedstock (71 MMGPY with ethylene feedstock and 9 MMGPY with biomass feedstock because 50 MMGPY are already produced by the existing ethylene plant). Clearly, one is interested more in the overall return on investment and, consequently, the optimal system is the 30 MMGPY biorefinery – 50 MMGPY ethylene-based plant integrated facility.

Then in Fig. 5.14, we use the results of the heat exchanger network calculation compared to the previous result that used the linear interpolation method. The difference is not significant. This suggests that linear interpolation for the cost of the HEN is satisfactory for conceptual design purposes.

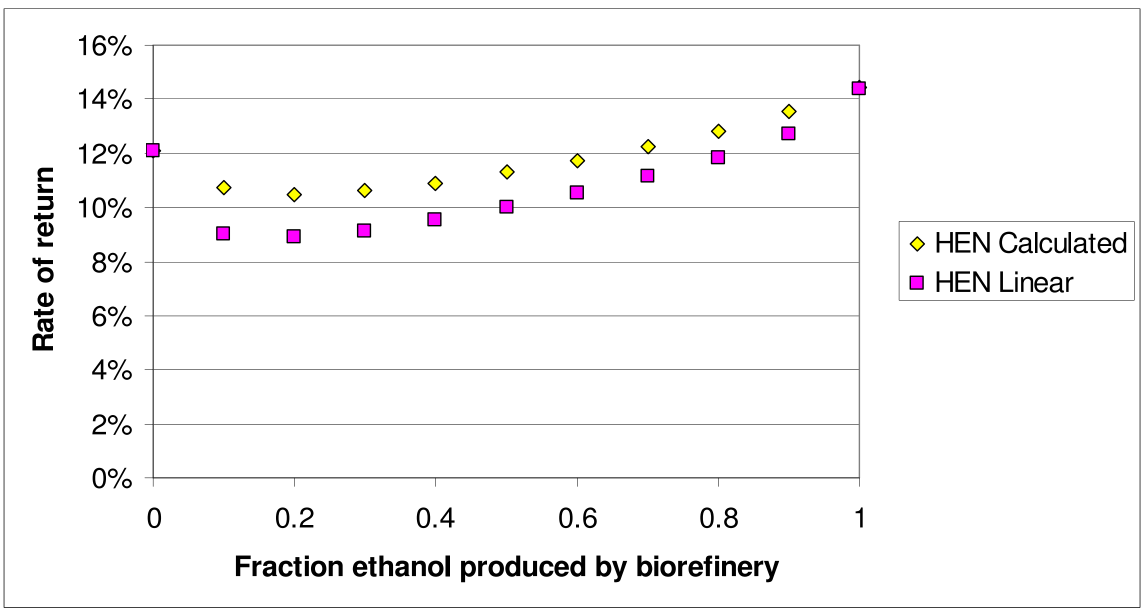

Figure 5.14 Rate of return with HEN calculated and HEN linear interpolation

CHAPTER VI

CONCLUSIONS AND RECOMMENDATIONS

This work has introduced a systematic approach to the retrofitting of existing facilities with an integrated bio-refinery. A hierarchical design procedure was developed to sequence the retrofitting activities. The proposed hierarchy involves internal process modification, operating-condition adjustment, and feedstock substitution. Next, new units are added followed by the incorporation of new production lines. Heat and mass integration techniques are used to link the units and streams. Ethanol production from two possible routes (fossil and biomass) was considered as an illustrating case study. The following specific conclusions were observed:

- The hierarchical design approach is computationally effective because it invokes the right level of details and calculations at each stage.
- Heat and mass integration can lead to cost-effective retrofitting of existing facilities.
- Integrating a new biorefinery with an adjacent facility improves the economics of both facilities.
- Biorefineries for ethanol production carry much promise in terms of supply and economics. Additionally, they provide significant reduction in GHG emissions that are associated with environmental and economic benefits.
- A standalone biorefinery for production of 30 MMGPY can be installed with a return on investment of about 10%. Upon integration with an

existing 50 MMGPY ethylene-based facility, the return on investment increases to about 14%. The key difference is heat integration which yields about \$6.5 MM/yr of cost savings.

- Although the biorefinery is economically more attractive than an ethylene-based process in the US, the trend is reversed in the Middle East where ethylene is cheaper and more readily available than biomass.

In this section, the study of the behavior of rate of return versus feedstock price and the operating cost shows that the recommended design depends on several key cost factors.

In the case of U.S., the curve shows that the bio refinery is the optimal option, but it may vary when the price changes for the utilities. After a change of the utilities, the fossil-refinery seems to be a better investment than bio refinery; this shows the sensitivity of U.S. market. In the Middle East case, the fossil refinery is the optimal option.

The closeness of the rate of return on U.S. and utilities case raised the issue with the linear interpolation of the heat duty. The calculation at each step for heat exchanger network shows that heat exchanger network technique could give a better integration for a different set of conditions. This improvement shows a little change in mater of rate of return on the mixture range. This improvement was not enough to change the optimal option in the U.S. case.

6.1 Recommendations for future work

This work constitutes the basis for several future activities to design and optimize several types of bio-refineries. The following are recommended topics to be investigated:

- Study of products other than ethanol (e.g., bio-diesel).

- Partial or complete substitution of the fossil-based feedstock with a biomass in the existing fossil-based process
- Scheduling studies for the use of a fossil feedstock a fraction of the year while using a biomass feedstock for the rest of the year depending on biomass availability. This work will involve the design and retrofitting of a multi-purpose process with multiple feedstocks and multiple products.
- Development of a computer-aided tool to automate the network synthesis, process integration, techno-economic analysis, and selection of optimum design alternatives.
- Incorporation of the bio-refinery in a life cycle analysis to evaluate the green house gas emissions for the whole cycle and the associated economic aspects for subsidy or mitigation.

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APPENDIX A

Table A1 Stream Table for ethylene plant for 30 MMGPY

	1	2	3	4	5	6	7	8	9	10	11	12	13
Temperature F	60	160.8	165.6	544.6	513	513	80	145	298.6	144	144	248.9	80
Pressure psi	255	1000	1000	1000	40	1000	1000	1000	1000	20	992	992	15
Vapor Frac	1	1	1	0	1	1	0	0.866	0	0	0	0	0.024
Mass Flow lb/hr	18079.28	352625.4	370704.6	107205.3	478212.5	478212.5	4013.804	378939.1	103287.1	30866.25	54441.22	157728.4	6591.263
Density lb/cuft	1.266	3.929	3.909	42.305	0.091	2.342	61.923	4.53	51.018	59.681	57.529	53.237	1.95
Mole Flow lbmol/hr													
ACETA-01					1	1		0.999	0.001			0.001	trace
DIETH-01					1	1		0.999	0.001			0.001	trace
ETHANOL					1	528.144		168.744	359.4	2.4	171	530.4	0.168
ETHYLENE	633.3	11346.7	11980		11981	11453.86		11442.4	11.454		1	12.454	11.496
METHANE	19.5	2094.5	2114		2115	2115		2112.885	2.115		1	3.115	0.003
N-BUTANO					1	1		0.999	0.001		1	1.001	<0.001
HYDROGEN					1	1			1			1	0.999
NAOH													
WATER		39.2	39.2	5950.8	5991	5463.856	222.8	892.256	4794.4	1707.2	2578.1	7372.5	347.422

	14	15	18	20	21	22	23	24	25	26	27	28
Temperature F	194.3	146	170.8	115	115	212	115	189.9	224.5	180	182.1	218.1
Pressure psi	15	992	20	35	35	20	35	40	40	18	18	18
Vapor Frac	0	1	0	1	1	1	0	0	<0.001	1	0	<0.001
Mass Flow lb/hr	143311.1	291628.7	31363.14	7.479	30.911	31370.62	31339.71	26.017	31084.33	1039.557	25807.3	5277.031
Density lb/cuft	56.125	3.952	48.331	0.011	0.041	0.094	50.716	45.119	45.954	0.124	46.174	56.722
Mole Flow lbmol/hr												
ACETA-01	< 0.001	0.026	0.958		0.003	0.958	0.955	0.034	< 0.001	1	< 0.001	trace
DIETH-01	< 0.001	0.103	0.822		0.021	0.822	0.801	0.02	trace	1	trace	trace
ETHANOL	362.035	0.144	524.453		0.292	524.453	524.161	0.47	521.43	20	519.466	1.964
ETHYLENE	0.957	9232.507	0.518		0.064	0.518	0.454	0.001	trace			
METHANE	0.07	2008.376	0.73		0.287	0.73	0.443	< 0.001	trace			
N-BUTANO	0.618	trace	0.117		<0.001	0.117	0.117	trace	0.117		trace	0.117
HYDROGEN	0.001		<0.001	3.71	3.533	3.71	0.177	trace	trace			
NAOH												
WATER	7025.078	21.476	392.111		0.123	392.111	391.988	0.075	391.549		104.134	287.415

Table A2 Distribution of hot and cold stream for pinch analysis

Cold Stream	T in	T out	Heat Duty	Specific Heat
Sc1	180	415	75.00	0.32
Sc2	144	226	2.50	0.03
Sc3	217	226	109.00	12.11
Sc4	166	212	1.50	0.03
Sc5	224	225	6.60	6.60
Sc6	200	218	99.00	5.50

Hot Stream	T in	T out	Heat Duty	Specific Heat
Sh1	541	300	75.00	0.31
Sh2	300	145	52.80	0.34
Sh3	195	80	12.60	0.11
Sh4	193	170	107.00	4.65
Sh5	212	115	2.40	0.02
Sh6	217	190	3.40	0.13
Sh7	182	100	1.60	0.02
Sh8	183	182	100.00	100.00
Sh9	218	135	0.38	0.00

Table A3 Composite interval diagram

Sh1	Sh2	Sh3	Sh4	Sh5	Sh6	Sh7	Sh8	Sh9	Heat	Cascade	Th	Tb	Tc	Sc1	Sc2	Sc3	Sc4	Sc5	Sc6	Heat	Cascade	Net Heat	Cascade Net Heat
											541	532	523										**181.51**
0.31									33.61	33.61	433	424	415								-	33.61	215.12
0.31									41.39	75.00	300	291	282	0.32						42.45	42.45	32.55	214.06
	0.34								19.08	94.08	244	235	226	0.32						17.87	60.32	33.76	215.26
	0.34								0.34	94.42	243	234	225	0.32	0.03	12.11		6.60		19.06	79.38	15.04	196.54
	0.34								0.34	94.76	242	233	224	0.32	0.03	12.11				12.46	91.84	2.92	184.42
	0.34								2.04	96.80	236	227	218	0.32	0.03	12.11				74.76	166.61	(69.80)	111.70
	0.34								0.34	97.14	235	226	217	0.32	0.03	12.11			5.50	17.96	184.57	(87.42)	94.08
	0.34								1.70	98.85	230	221	212	0.32	0.03				5.50	29.25	213.81	(114.97)	66.54
	0.34								4.09	102.93	218	209	200	0.32	0.03		0.03		5.50	70.59	284.40	(181.47)	0.04
	0.34							0.005	0.35	103.28	217	208	199	0.32	0.03		0.03			0.38	284.78	(181.51)	-
	0.34				0.13			0.005	2.36	105.63	212	203	194	0.32	0.03		0.03			1.91	286.69	(181.06)	0.44
	0.34			0.02	0.13			0.005	6.94	112.58	198	189	180	0.32	0.03		0.03			5.35	292.05	(179.47)	2.04
	0.34			0.02	0.13			0.005	1.49	114.06	195	186	177		0.03		0.03			0.19	292.24	(178.17)	3.33
	0.34	0.11		0.02	0.13			0.005	1.21	115.27	193	184	175		0.03		0.03			0.13	292.36	(177.09)	4.42
	0.34	0.11	4.65	0.02	0.13			0.005	15.77	131.05	190	181	172		0.03		0.03			0.19	292.55	(161.50)	20.00
	0.34	0.11	4.65	0.02				0.005	30.79	161.84	184	175	166		0.03		0.03			0.38	292.93	(131.09)	50.41
	0.34	0.11	4.65	0.02				0.005	5.13	166.97	183	174	165		0.03					0.03	292.96	(125.99)	55.52
	0.34	0.11	4.65	0.02			100	0.005	105.13	272.10	182	173	164		0.03					0.03	292.99	(20.89)	160.62
	0.34	0.11	4.65	0.02		0.02		0.005	61.81	333.92	170	161	152		0.03					0.37	293.36	40.56	222.07
	0.34	0.11		0.02		0.02		0.005	3.99	337.91	162	153	144		0.03					0.24	293.60	44.31	225.81
	0.34	0.11		0.02		0.02		0.005	8.48	346.39	145	136	127							-	293.60	52.79	234.30
		0.11		0.02		0.02		0.005	1.58	347.98	135	126	117							-	293.60	54.38	235.88
		0.11		0.02		0.02			3.08	351.05	115	106	97							-	293.60	57.45	238.96
		0.11				0.02			1.94	352.99	100	91	82							-	293.60	59.39	240.89
		0.11							2.19	355.18	80	71	62							-	293.60	61.58	**243.09**

Table A4 Equipment cost for reactor, column, and pumps of ethanol ethylene 50 MMGPY and 30 MMGPY

CASE:	Ethanol @ 2.81$/gal	Actual Plant 30 MMGPY
Ethylene		
	Plant A	30 MMGPY
	Plant B	50 MMGPY

Reactor, Column, and pumps

Number	Quantity	I.D. In	Height ft	Heat Duty	Price Each	Total Price
R-101	3	98.5	34		$ 459,578	$1,378,734
R-201	3	38	16		$ 117,697	$ 353,091
C-101	3	42	11		$ 121,054	$ 363,162
C-201	1	108	48		$ 557,370	$ 557,370
C-202	1	33	81		$ 347,216	$ 347,216
C-203	1	110	120		$1,524,867	$1,524,867
K-101	3					
K-102	3					
K-201	1					
F-101	3			12.7	$1,602,633	$4,807,899
Total						$9,332,339

Table A5 Equipment cost for heat exchanger and vessels of ethanol ethylene 50 MMGPY and 30 MMGPY

Heat Exchanger

Number	Quantity	Area (ft2)	Tin	Tout	Heat Duty Each	Heat Duty Total	Price Each	Total Price
E-101	3	1200	541	300	25	75	$ 143,481	$ 430,443
E-102	3	1200	180	415	25	75	$ 143,481	$ 430,443
E-103	3	550	300	145	17.6	52.8	$ 66,910	$ 200,730
E-104	1	190	144	226	2.5	2.5	$ 17,777	$ 17,777
E-200	1	240	195	80	12.6	12.6	$ 19,551	$ 19,551
E-201	1	3600	217	226	109	109	$ 98,635	$ 98,635
E-202	1	4500	193	170	107	107	$ 115,845	$ 115,845
E-203	1	50	166	212	1.5	1.5	$ 10,477	$ 10,477
E-204	1	260	212	115	2.4	2.4	$ 20,395	$ 20,395
E-205	1	390	224	225	6.6	6.6	$ 25,043	$ 25,043
E-206	1	170	217	190	3.4	3.4	$ 16,837	$ 16,837
E-207	1	3300	200	218	99	99	$ 92,951	$ 92,951
E-208	1	205	182	100	1.6	1.6	$ 18,232	$ 18,232
E-209	1	5100	183	182	100	100	$ 127,002	$ 127,002
E-210	1	160	218	135	0.38	0.38	$ 16,352	$ 16,352
Total								$1,640,713

Vessels and tanks

Number	Quantity	Volume (gal)	Price Each	Total Price
V-101	3	150	$ 11,701	$ 35,103
V-200	1	12000	$ 95,985	$ 95,985
V-201	1	7500	$ 68,242	$ 68,242
V-202	1	150	$ 11,733	$ 11,733
V-203	1	600	$ 23,596	$ 23,596
V-204	1	6000	$ 63,194	$ 63,194
V-205	1	20	$ 3,352	$ 3,352
T-201	1	3700	$ 52,279	$ 52,279
T-202	2	40000	$ 221,779	$ 443,558
Total				$ 797,042

Table A6 Summary of total capital investment

	Plant A	Plant B
Reactor and Columns	$ 9,332,339	$ 12,679,431
Heat Exchanger	$ 1,640,713	$ 2,229,163
Vessel	$ 797,042	$ 1,082,905
Total Equipments	$ 11,770,094	$ 15,991,499
Total Direct Plant Cost	$ 40,018,320	$ 54,371,097
Fixed Capital investment	$ 56,143,348	$ 76,279,451
working Capital	$ 9,416,075	$ 12,793,199
Total Capital Investment	$ 66,265,629	$ 90,032,141

Table A7 Operating cost

					Plant A	Plant B
	lb/hr		lb/yr	Cost $/lb		
Raw materials						
Ethylene	17764		149217600	0.39	$58,194,864	$ 96,991,440
Utilities	MM Btu/hr	MM Btu/yr	lb/yr			
Cooling water	283	2,374,512	118,725,600,000	0.00005	$16,621,584	$ 27,702,640
Steam	216	1,815,240	3,025,400,000	0.002	$10,891,440	$ 18,152,400
	gpm					
Process Water	802	-	3,379,178,880	0.00005	$ 168,959	$ 281,598
Total					$85,876,847	$143,128,078

Table A8 Sales

Sales		Plant A	Plant B
Ethanol	$/gal 2.81	$ 84,300,000	$ 140,500,000
Ether	$/lb 0.65	$ 2,095,510	$ 3,492,517
Total		$ 86,395,510	$ 143,992,517

Table A9 Rate of return calculations

year	Product Cost	Revenue	Income BT	Depreciation ratio	Depreciation	Income taxable	Income AT	Present Value
0	(66,265,629)							
1	$ 81,852,239	$ 86,395,510	$ 4,543,271	5%	$ 3,313,281	$ 1,229,990	$ 4,051,275	$ 3,950,296
2	$ 81,852,239	$ 86,395,510	$ 4,543,271	9.50%	$ 6,295,235	$ -	$ 4,543,271	$ 4,319,608
3	$ 81,852,239	$ 86,395,510	$ 4,543,271	8.55%	$ 5,665,711	$ -	$ 4,543,271	$ 4,211,940
4	$ 81,852,239	$ 86,395,510	$ 4,543,271	7.69%	$ 5,095,827	$ -	$ 4,543,271	$ 4,106,956
5	$ 81,852,239	$ 86,395,510	$ 4,543,271	6.93%	$ 4,592,208	$ -	$ 4,543,271	$ 4,004,589
6	$ 81,852,239	$ 86,395,510	$ 4,543,271	6.23%	$ 4,128,349	$ 414,923	$ 4,377,302	$ 3,762,128
7	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,595,052
8	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,505,444
9	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,418,069
10	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,332,872
11	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,249,799
12	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,168,797
13	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,089,813
14	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 3,012,798
15	$ 81,852,239	$ 86,395,510	$ 4,543,271	5.90%	$ 3,909,672	$ 633,599	$ 4,289,832	$ 2,937,703
16	$ 81,852,239	$ 86,395,510	$ 4,543,271	3.00%	$ 1,987,969	$ 2,555,303	$ 3,521,150	$ 2,351,203
17	$ 81,852,239	$ 86,395,510	$ 4,543,271	0.00%	$ -	$ 4,543,271	$ 2,725,963	$ 1,774,857
18	$ 81,852,239	$ 86,395,510	$ 4,543,271	0.00%	$ -	$ 4,543,271	$ 2,725,963	$ 1,730,618
19	$ 81,852,239	$ 86,395,510	$ 4,543,271	0.00%	$ -	$ 4,543,271	$ 2,725,963	$ 1,687,481
20	$ 81,852,239	$ 95,811,586	$ 13,959,347	0.00%	$ -	$ 13,959,347	$ 8,375,608	$ 5,055,606
								$ 66,265,629

ROR	**2.56%**

APPENDIX B

Table B1 Streams table (1of 3) for 69 MMGPY

	1	2	3	4	5	6	7	8	9	10	11	12	13
Temperature F	113	68	211.5	221.1	122	141.8	149		175.2	182.7	140	140	251.1
Pressure psi	14.7	49.97	14.7	14.7	14.7	14.7	14.7		13.23	14.7	29.39	29.39	29.39
Vapor Frac	0	0	0.077	0	0	0.001	0		0	1	1	0	0
Mass Flow lb/hr	143104.3	4638.526	116016.4	407078.2	87831.97	605531.4	894500.6		903307.2	70380.53	1170.13	133826.1	768311
Volume Flow cuft/hr	2320.315	40.938	241673	6444.113	1394.269	18603.13	14027.56		15198.82	1.06E+06	6092.81	2446.859	15041.55
Mass Flow lb/hr													
ETHANOL	49		92	113	4	149	106		54,829	3,411	57	49,904	4,869
WATER	99,223		114,263	303,242	81,180	500,754	747,692		806,749	20,861	39	81,331	725,379
GLUCOSE				6,417	1,909	4,534	73,412		331				331
XYLOSE			trace	42,917	2,873	40,335	40,094		2,186				2,186
SOLUNKN	16,352			16,352	1,094	18,338	21,121		21,121				21,121
CSL	90		105	279	18	461	593		3,294	94	0	332	2,962
ACETIC	511		1,310	6,243	425	7,682	9,830		9,862	163	<0.001	0	9,861
SULFURIC		4,639	0	4,638	311	4,376	340		339	<0.001			339
FURFURAL	75		4	71	18	472	1,250		1,264	5	trace	trace	1,264
CO2				2		31	64		3,333	45,846	1,074	2,259	trace
CH4													
O2													
N2													
CELLULOS				62,980	62,989		3,150		3,135				3,135
XYLAN				968	970		972		972				972
CELLULAS					112		1,431		1,431				1,431
BIOMASS													
ZYMO					190		300		695				695
LIGNIN				33,074	33,173		33,230		33,049				33,049
GYPSUM							62		62				62
SOLSLDS	26803.8			26,804		26,804							

Table B2 Streams table (2of 3) for 69 MMGPY

	14	15	16	17	18	19	20	21	22	23	24	25	26
Temperature F	219.9	219.9	236.5	189.2	189.2	186.4	186.4	160.5	178	160.5	147.6		
Pressure psi	24.98	24.98	24.98	8.82	8.82	47.03	47.03	4.56	8.82	4.56	3.09		
Vapor Frac	1	0.785	0	1	0	0	0	0	1	1	1		
Mass Flow lb/hr	2312.004	51718.89	78617.91	148939.2	630080.9	39253.81	590352.6	229884.3	148939.2	152247.7	178058.7	301,187	
Volume Flow cuft/hr	15773.36	352222.3	1421.19	6.41E+06	10479.72	653.074	9806.057	3756.683	6.41E+06	1.22E+07	2.05E+07		
Mass Flow lb/hr													
ETHANOL		49,860	40	1,722	3,146	189	2,957	877	1,722	905	1,345	2,627	3,972
WATER		3,032	78,299	145,377	590,711	35,443	554,796	205,767	145,377	149,422	173,490	294,799	468,289
GLUCOSE					331	33	298	126					
XYLOSE				trace	2,186	219	1,967	1,387	trace	trace	trace	trace	trace
SOLUNKN					21,121	2,112	19,009	14,612					
CSL	53		279	585	2,377	238	2,137	259	585	167	243	752	995
ACETIC		< 0.001	0	1,216	8,645	865	7,781	6,620	1,216	1,748	2,979	2,964	5,943
SULFURIC				0	339	34	305	236	0	0	0	0	0
FURFURAL				39	1,225	123	1,103		39	7	1	46	47
CO2	2,259		trace										
CH4													
O2													
N2													
CELLULOS					3,135	3,135		44					
XYLAN					972	972		15					
CELLULAS					1,431	1,431							
BIOMASS													
ZYMO					695	695		899					
LIGNIN					33,049	33,049		472					
GYPSUM					62	62							
SOLSLDS													

Table B3 Streams table (3of 3) for 69 MMGPY

	27	28	29	30	31	32	33	34	35	36	37
Temperature F				105.8					166.5	174.7	147.6
Pressure psi				14.7					13.23	13.23	3.09
Vapor Frac				0					1	0	0
Mass Flow lb/hr				89450.06					66844.14	89198.68	28386.87
Volume Flow cuft/hr				1385.805					1.02E+06	1489.89	459.938
Mass Flow lb/hr											
ETHANOL				11					87	3,535	436
WATER				74,769					15,235	85,006	11,407
GLUCOSE				7,341							126
XYLOSE				4,009							1,387
SOLUNKN				2,112							14,612
CSL				59					0	94	183
ACETIC				983					0	163	
SULFURIC				34					trace	<0.001	236
FURFURAL				125					trace	5	
CO2				6					51,452	395	
CH4											
O2									70	0	
N2											
CELLULOS				315							44
XYLAN				97							15
CELLULAS				143							
BIOMASS											
ZYMO				30							899
LIGNIN				3,323							472
GYPSUM				6							
SOLSLDS											

Table B4 Distribution of hot and cold steam

Cold stream	T in	T out	Heat duty	Specific heat
Sh10	235	140	0.63	0.01
Sh11	214	86	45.09	0.35
Sh12	131	122	4.21	0.47
Sh13	149	106	8.44	0.20
Sh14	138	95	3.87	0.09
Sh15	149	145	91.85	22.96

Hot stream	T in	T out	Heat duty	Specific heat
Sc9	124	149	14.53	0.58
Sc8	106	203	38.83	0.40
Sc7	235	253	58.26	3.24
Sc10	203	212	3.76	0.42
Sc11	171	189	58.85	3.27
Sc12	142	160	95.76	5.32
Sc13	130	147	91.54	5.38

Table B5Composite interval diagram

Sh10	Sh11	Sh12	Sh13	Sh14	Heat Available	Cascade	Th	Tb	Tc	Sc7	Sc8	Sc9	Sc10	Heat Available	Cascade	Net Heat	Cascade Net Heat
							271	262	253					-	-	-	80.81
							253	244	235	3.24				58.26	58.26	(58.26)	22.55
						-	235	226	217					-	58.26	(58.26)	22.55
0.01					0.03	0.03	230	221	212					-	58.26	(58.23)	22.58
0.01					0.06	0.09	221	212	203				0.42	3.76	62.02	(61.93)	18.88
0.01					0.05	0.14	214	205	196		0.40			2.80	64.82	(64.68)	16.13
0.01	0.35				16.87	17.01	167	158	149		0.40			18.81	83.64	(66.63)	14.19
0.01	0.35				6.46	23.47	149	140	131		0.40	0.58		17.66	101.30	(77.83)	2.98
0.01	0.35		0.20		3.89	27.36	142	133	124		0.40	0.58		6.87	108.17	(80.81)	-
0.01	0.35		0.20		1.11	28.47	140	131	122		0.40			0.80	108.97	(80.50)	0.31
	0.35		0.20		1.10	29.56	138	129	120		0.40			0.80	109.77	(80.21)	0.61
	0.35		0.20	0.09	4.47	34.03	131	122	113		0.40			2.80	112.57	(78.54)	2.27
	0.35	0.47	0.20	0.09	7.74	41.78	124	115	106		0.40			2.80	115.37	(73.60)	7.21
	0.35	0.47	0.20	0.09	2.21	43.99	122	113	104					-	115.37	(71.39)	9.43
	0.35		0.20	0.09	10.22	54.20	106	97	88					-	115.37	(61.17)	19.64
	0.35			0.09	4.86	59.07	95	86	77					-	115.37	(56.30)	24.51
	0.35				3.17	62.24	86	77	68					-	115.37	(53.13)	27.68

Table B6 Summary of total capital investment for Biomass 69 and 30 MMGPY, respectively plant A and B

	Plant A	Plant B
A100	$ 4,848,032	$ 2,941,228
A200	$ 16,965,825	$ 10,292,913
A300	$ 8,180,970	$ 4,963,272
A400	$ -	$ -
A500	$ 11,934,417	$ 7,240,432
A600	$ 4,312,647	$ 2,616,419
A700	$ 1,426,519	$ 865,448
A800	$ -	$ -
A900	$ 351,687	$ 213,363
Total equipments	$ 48,020,097	$ 29,133,075
Total direct plant cost	$ 137,817,679	$ 83,611,926
Fixed capital investment	$ 195,441,795	$ 118,571,617
Working capital	$ 32,653,666	$ 19,810,491
Total cost investment	$ 230,976,667	$ 140,130,093

Table B7 Operating cost

	Plant A	Plant B
Feedstock	$ 14,500,000	$ 6,304,348
CSL	$ 2,375,000	$ 1,032,609
Cellulase	$ 8,750,000	$ 3,804,348
Waste dis.	$ 2,500,000	$ 1,086,957
Cooling water	$ 5,678,904	$ 2,469,089
Steam	$ 4,112,640	$ 1,788,104
Total cost	$ 40,875,000	$ 16,485,454

Table B8 Sales

Sales		
Ethanol	1.2 $/gal	$ 36,000,000
Syrup		$ 2,758,621
Total		$ 38,758,621

Table B9 Rate of return calculations

Year	Product cost	Revenue	Income before tax	Depreciation ratio	Depreciation	Income taxable	Income after tax	Present value
0	$(140,130,093)							
1	$ 16,485,454	$ 38,758,621	$ 22,273,167	5%	$ 5,928,581	$ 16,344,586	$15,735,332	$ 14,307,330
2	$ 16,485,454	$ 38,758,621	$ 22,273,167	10%	$ 11,264,304	$ 11,008,863	$17,869,622	$ 14,773,408
3	$ 16,485,454	$ 38,758,621	$ 22,273,167	9%	$ 10,137,873	$ 12,135,294	$17,419,049	$ 13,094,003
4	$ 16,485,454	$ 38,758,621	$ 22,273,167	8%	$ 9,118,157	$ 13,155,009	$17,011,163	$ 11,626,919
5	$ 16,485,454	$ 38,758,621	$ 22,273,167	7%	$ 8,217,013	$ 14,056,154	$16,650,705	$ 10,347,750
6	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 7,387,012	$ 14,886,155	$16,318,705	$ 9,221,077
7	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 8,303,838
8	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 7,550,253
9	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 6,865,057
10	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 6,242,044
11	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 5,675,570
12	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 5,160,505
13	$ 16,485,454	$ 38,758,621	$ 22,273,167	5%	$ 5,928,581	$ 16,344,586	$15,735,332	$ 4,568,257
14	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 4,266,360
15	$ 16,485,454	$ 38,758,621	$ 22,273,167	6%	$ 6,995,725	$ 15,277,441	$16,162,190	$ 3,879,182
16	$ 16,485,454	$ 38,758,621	$ 22,273,167	3%	$ 3,557,149	$ 18,716,018	$14,786,759	$ 3,226,975
17	$ 16,485,454	$ 38,758,621	$ 22,273,167			$ 22,273,167	$13,363,900	$ 2,651,786
18	$ 16,485,454	$ 38,758,621	$ 22,273,167			$ 22,273,167	$13,363,900	$ 2,411,133
19	$ 16,485,454	$ 38,758,621	$ 22,273,167			$ 22,273,167	$13,363,900	$ 2,192,319
20	$ 16,485,454	$ 58,569,112	$ 42,083,658			$ 42,083,658	$25,250,195	$ 3,766,327
								$ 140,130,093

ROR	9.98%

APPENDIX C

Table C1 Economics for 80 MMGPY fossil/fossil hybrid plant

Economics for 80 MMGPY Ethylene Plant

Total Capital Investment	$	66,265,629
Operating Cost	$	218,272,637
Income	$	230,388,028
Net Revenue	$	7,269,234

Table C2 Rate of return calculations

year	Product Cost	Revenue	Income before tax	Depreciation ratio	Depreciation	Income taxable	Income after tax	Present Value
0	$(66,265,629)							
1	$218,272,637	$ 230,388,028	$ 12,115,390	5%	$ 3,313,281	$ 8,802,109	$ 8,594,547	$ 7,657,730
2	$218,272,637	$ 230,388,028	$ 12,115,390	9.50%	$ 6,295,235	$ 5,820,156	$ 9,787,328	$ 7,769,950
3	$218,272,637	$ 230,388,028	$ 12,115,390	8.55%	$ 5,665,711	$ 6,449,679	$ 9,535,519	$ 6,744,899
4	$218,272,637	$ 230,388,028	$ 12,115,390	7.69%	$ 5,095,827	$ 7,019,564	$ 9,307,565	$ 5,866,030
5	$218,272,637	$ 230,388,028	$ 12,115,390	6.93%	$ 4,592,208	$ 7,523,182	$ 9,106,117	$ 5,113,502
6	$218,272,637	$ 230,388,028	$ 12,115,390	6.23%	$ 4,128,349	$ 7,987,042	$ 8,920,574	$ 4,463,289
7	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 3,937,790
8	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 3,508,566
9	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 3,126,127
10	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 2,785,375
11	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 2,481,766
12	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 2,211,250
13	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 1,970,221
14	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 1,755,464
15	$218,272,637	$ 230,388,028	$ 12,115,390	5.90%	$ 3,909,672	$ 8,205,718	$ 8,833,103	$ 1,564,116
16	$218,272,637	$ 230,388,028	$ 12,115,390	3.00%	$ 1,987,969	$ 10,127,422	$ 8,064,422	$ 1,272,348
17	$218,272,637	$ 230,388,028	$ 12,115,390	0.00%	$ -	$ 12,115,390	$ 7,269,234	$ 1,021,876
18	$218,272,637	$ 230,388,028	$ 12,115,390	0.00%	$ -	$ 12,115,390	$ 7,269,234	$ 910,491
19	$218,272,637	$ 230,388,028	$ 12,115,390	0.00%	$ -	$ 12,115,390	$ 7,269,234	$ 811,246
20	$218,272,637	$ 239,804,103	$ 21,531,466	0.00%	$ -	$ 21,531,466	$ 12,918,879	$ 1,284,593
								$ 66,256,629

ROR	**12.23%**

Table C 3 Economics for 80 MMGPY Hybrid bio/fossil Plant

Economics for 80 MMGPY hybrid plant

Total cost investment	$ 140,130,093
Operating cost	$ 152,905,852
Sales	$ 182,751,138
Net revenue	$ 17,907,171

Table C4 Rate of return calculations

Year	Product Cost	Revenue	Income BT	Depreciation ratio	Depreciation	Income taxable	Income AT	Present Value
0	$(140,130,093)							
1	$ 152,905,852	$182,751,138	$ 29,845,286	5.00%	$ 5,928,581	$ 23,916,705	$20,278,604	$ 17,801,666
2	$ 152,905,852	$182,751,138	$ 29,845,286	9.50%	$ 11,264,304	$ 18,580,982	$22,412,893	$ 17,272,019
3	$ 152,905,852	$182,751,138	$ 29,845,286	8.55%	$ 10,137,873	$ 19,707,413	$21,962,321	$ 14,857,509
4	$ 152,905,852	$182,751,138	$ 29,845,286	7.69%	$ 9,118,157	$ 20,727,128	$21,554,434	$ 12,800,502
5	$ 152,905,852	$182,751,138	$ 29,845,286	6.93%	$ 8,217,013	$ 21,628,273	$21,193,977	$ 11,049,062
6	$ 152,905,852	$182,751,138	$ 29,845,286	6.23%	$ 7,387,012	$ 22,458,274	$20,861,976	$ 9,547,529
7	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 8,318,462
8	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 7,302,400
9	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 6,410,446
10	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 5,627,439
11	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 4,940,074
12	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 4,336,666
13	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 3,806,963
14	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 3,341,960
15	$ 152,905,852	$182,751,138	$ 29,845,286	5.90%	$ 6,995,725	$ 22,849,560	$20,705,462	$ 2,933,755
16	$ 152,905,852	$182,751,138	$ 29,845,286	3.00%	$ 3,557,149	$ 26,288,137	$19,330,031	$ 2,404,330
17	$ 152,905,852	$182,751,138	$ 29,845,286			$ 29,845,286	$17,907,171	$ 1,955,289
18	$ 152,905,852	$182,751,138	$ 29,845,286			$ 29,845,286	$17,907,171	$ 1,716,460
19	$ 152,905,852	$182,751,138	$ 29,845,286			$ 29,845,286	$17,907,171	$ 1,506,802
20	$ 152,905,852	$202,561,629	$ 49,655,777			$ 49,655,777	$29,793,466	$ 2,200,761

ROR	13.91%

Table C5 Distribution of hot stream for each step (1of2)

			0		**0.1**		**0.2**		**0.3**		**0.4**	
Hot stream	Tin	Tout	Heat Duty MMBtu/hr	Specific Heat MMBtu/hr/F								
Sh1	541	300	200	**0.83**	192.5	**0.80**	185	**0.77**	177.5	**0.74**	170	**0.71**
Sh10	235	140	-	**-**	0.06	**0.00**	0.13	**0.00**	0.2	**0.00**	0.25	**0.00**
Sh11	214	86	-	**-**	4.51	**0.04**	9.02	**0.07**	13.5	**0.11**	18.03	**0.14**
Sh12	131	122	-	**-**	0.42	**0.05**	0.84	**0.09**	1.2	**0.14**	1.68	**0.19**
Sh13	149	106	-	**-**	0.84	**0.02**	1.69	**0.04**	2.5	**0.06**	3.38	**0.08**
Sh14	138	95	-	**-**	0.39	**0.01**	0.77	**0.02**	1.16	**0.03**	1.55	**0.04**
Sh15	149	145	-	**-**	9.18	**2.30**	18.37	**4.59**	27.5	**6.89**	36.74	**9.18**
Sh2	300	145	140.8	**0.91**	135.5	**0.87**	130.2	**0.84**	124.9	**0.81**	119.6	**0.77**
Sh3	195	80	33.6	**0.29**	32.3	**0.28**	31.08	**0.27**	29.8	**0.26**	28.56	**0.25**
Sh4	193	170	285.3	**12.41**	274	**11.94**	264	**11.48**	253	**11.01**	242	**10.54**
Sh5	212	115	6.4	**0.07**	6.16	**0.06**	5.92	**0.06**	5.68	**0.06**	5.44	**0.06**
Sh6	217	190	9.07	**0.34**	8.73	**0.32**	8.39	**0.31**	8.05	**0.30**	7.71	**0.29**
Sh7	182	100	4.27	**0.05**	4.27	**0.05**	4.27	**0.05**	4.27	**0.05**	4.27	**0.05**
Sh8	183	182	266.	266.	266.	266.	266.	266.	266.	266.	266.	266.
Sh9	218	135	1.	1.	1.	1.	1.	1.	1.	1.	1.	1.

Table C6 Distribution of hot stream for each step (2of2)

0.5		**0.6**		**0.7**		**0.8**		**0.9**		**1**	
162	**0.67**	155	**0.64**	147.5	**0.61**	140	**0.58**	132.50	**0.55**	125	**0.52**
0.3	**0.00**	0.38	**0.00**	0.4	**0.00**	0.51	**0.01**	0.57	**0.01**	0.6	**0.01**
22	**0.18**	27.05	**0.21**	31.5	**0.25**	36.07	**0.28**	40.58	**0.32**	45	**0.35**
2	**0.23**	2.53	**0.28**	2.9	**0.33**	3.37	**0.37**	3.79	**0.42**	4.2	**0.47**
4.	**0.10**	5.06	**0.12**	5.9	**0.14**	6.75	**0.16**	7.60	**0.18**	8.4	**0.20**
1.9	**0.04**	2.32	**0.05**	2.7	**0.06**	3.10	**0.07**	3.48	**0.08**	3.9	**0.09**
45.9	**11.48**	55.11	**13.78**	64.3	**16.07**	73.48	**18.37**	82.66	**20.67**	91.9	**22.96**
114	**0.74**	109	**0.70**	103.8	**0.67**	98.56	**0.64**	93.28	**0.60**	88	**0.57**
27.3	**0.24**	26.04	**0.23**	24.7	**0.22**	23.52	**0.20**	22.26	**0.19**	21	**0.18**
231	**10.08**	221	**9.61**	210	**9.15**	199.7	**8.68**	189	**8.22**	178	**7.75**
5.2	**0.05**	4.96	**0.05**	4.7	**0.05**	4.48	**0.05**	4.24	**0.04**	4	**0.04**
7.4	**0.27**	7.03	**0.26**	6.6	**0.25**	6.35	**0.24**	6.01	**0.22**	5.7	**0.21**
4.3	**0.05**	4.27	**0.05**	4.2	**0.05**	4.27	**0.05**	4.27	**0.05**	4.3	**0.05**
266.	266.	266.	266.	266.	266.	266.	266.	266.	266.	266.	266.
1.	1.	1.	1.	1.	1.	1.	1.	1.	1.	1.	1.

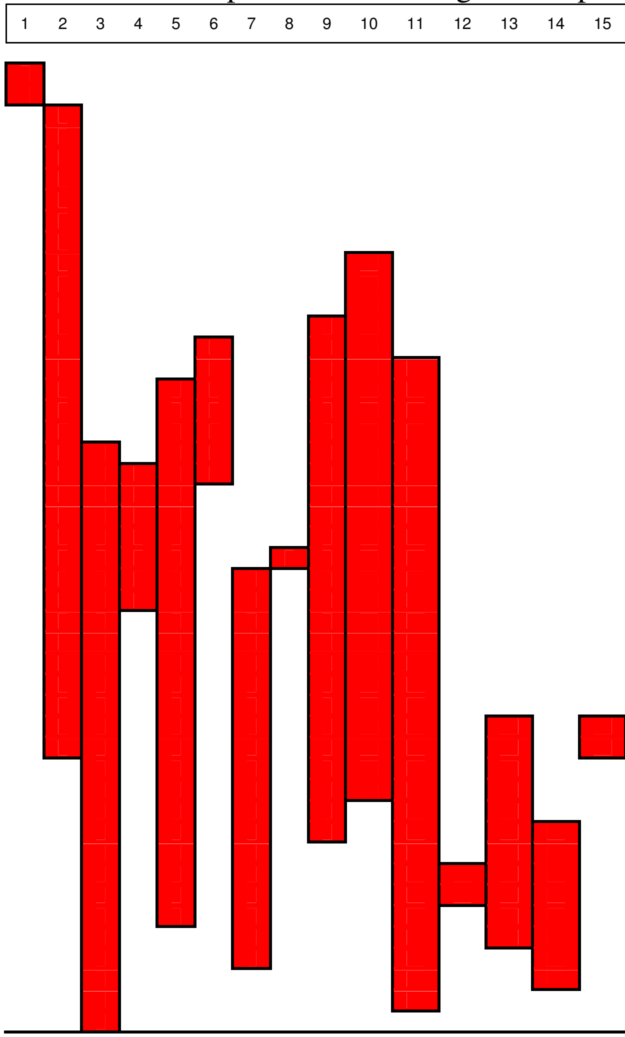
Table C7 Distribution of cold stream for each step (1of2)

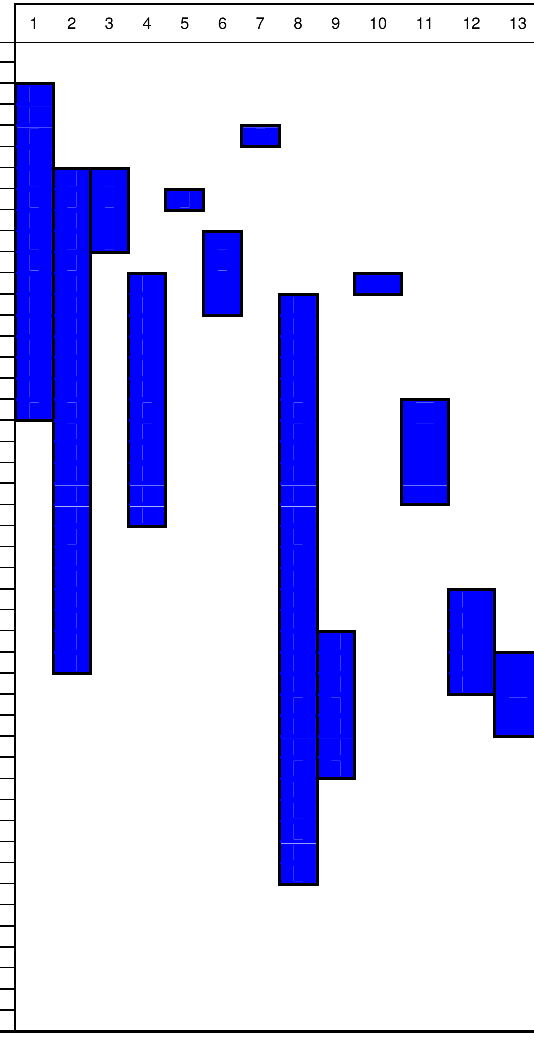
Hot Stream	Tin	Tout	Heat Duty MMBtu/hr	Specific Heat MMBtu/hr/F	0.1		0.2		0.3		0.4	
Sc1	180	415	200	**0.85**	192.50	**0.82**	185	**0.79**	177.50	**0.76**	170	**0.72**
Sc10	203	212	-	**-**	0.38	**0.04**	0.75	**0.08**	1.13	**0.13**	1.50	**0.17**
Sc11	171	189	-	**-**	5.88	**0.33**	11.77	**0.65**	17.65	**0.98**	23.54	**1.31**
Sc12	142	160	-	**-**	9.58	**0.53**	19.15	**1.06**	28.73	**1.60**	38.30	**2.13**
Sc13	130	147	-	**-**	9.15	**0.54**	18.31	**1.08**	27.46	**1.62**	36.62	**2.15**
Sc2	144	226	6.67	**0.08**	6.42	**0.08**	6.17	**0.08**	5.92	**0.07**	5.67	**0.07**
Sc3	217	226	290.67	**32.30**	279.77	**31.09**	268.87	**29.87**	257.97	**28.66**	247.07	**27.45**
Sc4	166	212	4.00	**0.09**	3.85	**0.08**	3.70	**0.08**	3.55	**0.08**	3.40	**0.07**
Sc5	224	225	17.60	**17.60**	16.94	**16.94**	16.28	**16.28**	15.62	**15.62**	14.96	**14.96**
Sc6	200	218	264.00	**14.67**	264.00	**14.67**	264.00	**14.67**	264.00	**14.67**	264.00	**14.67**
Sc7	235	253	-	**-**	5.83	**0.32**	11.65	**0.65**	17.48	**0.97**	23.30	**1.29**
Sc8	106	203	-	**-**	3.88	**0.04**	7.77	**0.08**	11.65	**0.12**	15.53	**0.16**
Sc9	124	149	-	**-**	1.45	**0.06**	2.91	**0.12**	4.36	**0.17**	5.81	**0.23**

Table C8 Distribution of cold stream for each step (2of2)

0.5		0.6		0.7		0.8		0.9		1	
162.50	**0.69**	155	**0.66**	147.50	**0.63**	140	**0.60**	132.50	**0.56**	125.00	**0.53**
1.88	**0.21**	2.26	**0.25**	2.63	**0.29**	3.01	**0.33**	3.38	**0.38**	3.76	**0.42**
29.42	**1.63**	35.31	**1.96**	41.19	**2.29**	47.08	**2.62**	52.96	**2.94**	58.85	**3.27**
47.88	**2.66**	57.46	**3.19**	67.03	**3.72**	76.61	**4.26**	86.18	**4.79**	95.76	**5.32**
45.77	**2.69**	54.93	**3.23**	64.08	**3.77**	73.23	**4.31**	82.39	**4.85**	91.54	**5.38**
5.42	**0.07**	5.17	**0.06**	4.92	**0.06**	4.67	**0.06**	4.42	**0.05**	4.17	**0.05**
236.17	**26.24**	225.27	**25.03**	214.37	**23.82**	203.47	**22.61**	192.57	**21.40**	181.67	**20.19**
3.25	**0.07**	3.10	**0.07**	2.95	**0.06**	2.80	**0.06**	2.65	**0.06**	2.50	**0.05**
14.30	**14.30**	13.64	**13.64**	12.98	**12.98**	12.32	**12.32**	11.66	**11.66**	11.00	**11.00**
264.00	**14.67**	264.00	**14.67**	264.00	**14.67**	264.00	**14.67**	264.00	**14.67**	264.00	**14.67**
29.13	**1.62**	34.96	**1.94**	40.78	**2.27**	46.61	**2.59**	52.43	**2.91**	58.26	**3.24**
19.41	**0.20**	23.30	**0.24**	27.18	**0.28**	31.06	**0.32**	34.94	**0.36**	38.83	**0.40**
7.26	**0.29**	8.72	**0.35**	10.17	**0.41**	11.62	**0.46**	13.07	**0.52**	14.53	**0.58**

Table C9 Composite interval diagram template for X=0





Heat	Ce				Heat	C	Net Heat	C Net Heat
	0	541	532	523				484.01
89.63	89.63	433	424	415	-	0	89.63	573.64
110.37	200.00	300	291	282	113.19	113.19	86.81	570.82
26.34	226.34	271	262	253	24.68	137.87	88.47	572.48
16.35	242.69	253	244	235	15.32	153.19	89.50	573.52
8.18	250.87	244	235	226	7.66	160.85	90.02	574.03
0.91	251.78	243	234	225	33.23	194.08	57.70	541.71
0.91	252.69	242	233	224	50.83	244.91	7.78	491.79
5.45	258.14	236	227	218	199.37	444.28	(186.14)	297.87
0.91	259.05	235	226	217	47.90	492.18	(233.13)	250.88
4.54	263.59	230	221	212	78.00	570.17	(306.58)	177.43
8.18	271.76	221	212	203	141.17	711.34	(439.58)	44.43
2.73	274.49	218	209	200	47.06	758.40	(483.91)	0.10
0.92	275.41	217	208	199	1.02	759.42	(484.01)	-
3.77	279.18	214	205	196	3.06	762.48	(483.30)	0.71
2.51	281.69	212	203	194	2.04	764.52	(482.83)	1.19
6.61	288.30	207	198	189	5.10	769.62	(481.31)	2.70
11.90	300.20	198	189	180	9.17	778.79	(478.59)	5.43
3.97	304.17	195	186	177	0.50	779.29	(475.12)	8.89
3.23	307.40	193	184	175	0.34	779.63	(472.23)	11.78
42.06	349.46	190	181	172	0.50	780.14	(430.67)	53.34
13.68	363.15	189	180	171	0.17	780.30	(417.16)	66.86
68.42	431.57	184	175	166	0.84	781.14	(349.58)	134.44
13.68	445.25	183	174	165	0.08	781.23	(335.97)	148.04
280.35	725.60	182	173	164	0.08	781.31	(55.70)	428.31
54.95	780.55	178	169	160	0.33	781.63	(1.08)	482.93
109.89	890.44	170	161	152	0.65	782.28	108.16	592.17
3.99	894.44	167	158	149	0.24	782.53	111.91	595.92
2.66	897.10	165	156	147	0.16	782.69	114.41	598.42
3.99	901.09	162	153	144	0.24	782.93	118.16	602.17
2.66	903.75	160	151	142	-	782.93	120.82	604.83
14.64	918.39	149	140	131	-	782.93	135.46	619.47
1.33	919.72	148	139	130	-	782.93	136.79	620.80
3.99	923.71	145	136	127	-	782.93	140.78	624.79
1.27	924.98	142	133	124	-	782.93	142.05	626.06
0.84	925.82	140	131	122	-	782.93	142.89	626.90
0.84	926.67	138	129	120	-	782.93	143.74	627.75
1.27	927.94	135	126	117	-	782.93	145.00	629.02
1.64	929.58	131	122	113	-	782.93	146.64	630.66
2.87	932.45	124	115	106	-	782.93	149.52	633.53
0.82	933.27	122	113	104	-	782.93	150.34	634.35
2.87	936.14	115	106	97	-	782.93	153.21	637.22
3.10	939.24	106	97	88	-	782.93	156.30	640.32
2.07	941.30	100	91	82	-	782.93	158.37	642.38
1.46	942.76	95	86	77	-	782.93	159.83	643.84
2.63	945.39	86	77	68	-	782.93	162.46	646.47
1.75	947.15	80	71	62	-	782.93	164.21	648.23

Table C10 Net heat cascade for each step

Fraction	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
Tb											
532	484.01	482.11	481.43	482.85	485.48	488.79	492.10	495.42	498.73	502.04	505.35
424	573.64	568.38	564.33	562.39	561.66	561.61	561.56	561.51	561.47	561.42	561.37
291	570.82	565.67	561.72	559.89	559.27	559.32	559.38	559.44	559.49	559.55	559.61
262	572.48	567.27	563.26	561.37	560.68	560.67	560.67	560.66	560.66	560.65	560.65
244	573.52	562.43	552.56	544.80	538.25	532.38	526.51	520.64	514.77	508.90	503.03
235	574.03	562.93	553.04	545.26	538.69	532.80	526.91	521.02	515.13	509.24	503.35
234	541.71	531.82	523.15	516.58	511.22	506.54	501.86	497.19	492.51	487.83	483.15
233	491.79	483.77	476.97	472.27	468.79	465.98	463.17	460.37	457.56	454.76	451.95
227	297.87	297.12	297.59	300.17	303.95	308.42	312.89	317.35	321.82	326.28	330.75
226	250.88	251.35	253.03	256.82	261.81	267.49	273.17	278.85	284.53	290.21	295.88
221	177.43	177.90	179.59	183.39	188.39	194.08	199.76	205.45	211.14	216.82	222.51
212	44.43	44.57	45.93	49.39	54.06	59.42	64.77	70.12	75.48	80.83	86.18
209	0.10	0.13	1.38	4.74	9.31	14.56	19.80	25.05	30.30	35.55	40.80
208	-	-	1.21	4.54	9.07	14.28	19.49	24.71	29.92	35.13	40.35
205	0.71	0.57	1.64	4.82	9.21	14.28	19.34	24.41	29.48	34.55	39.62
203	1.19	1.02	2.06	5.22	9.58	14.62	19.67	24.71	29.76	34.80	39.85
198	2.70	2.46	3.43	6.51	10.79	15.76	20.73	25.70	30.67	35.64	40.61
189	5.43	2.11	-	-	1.21	3.10	4.99	6.88	8.78	10.67	12.56
186	8.89	4.45	1.22	0.10	0.19	0.96	1.73	2.50	3.26	4.03	4.80
184	11.78	6.57	2.57	0.68	-	-	-	-	-	-	-
181	53.34	45.58	39.03	34.59	31.36	28.81	26.26	23.70	21.15	18.60	16.05
180	66.86	58.26	50.87	45.59	41.52	38.13	34.75	31.36	27.97	24.58	21.20
175	134.44	123.28	113.35	105.51	98.89	92.95	87.01	81.07	75.13	69.19	63.25
174	148.04	136.37	125.92	117.58	110.44	103.99	97.53	91.08	84.62	78.17	71.72
173	428.31	416.13	405.16	396.30	388.66	381.69	374.72	367.75	360.78	353.82	346.85
169	482.93	468.70	455.67	444.76	435.06	426.03	417.01	407.99	398.96	389.94	380.92
161	592.17	569.57	548.18	528.90	510.83	493.44	476.05	458.66	441.27	423.88	406.49
158	595.92	571.58	548.45	527.43	507.61	488.48	469.35	450.22	431.09	411.96	392.83
156	598.42	572.80	548.39	526.09	505.00	484.59	464.19	443.78	423.37	402.96	382.55
153	602.17	573.02	545.08	519.25	494.63	470.69	446.74	422.80	398.86	374.92	350.98
151	604.83	573.32	543.02	514.83	487.85	461.54	435.24	408.94	382.64	356.34	330.04
140	619.47	580.83	543.40	508.07	473.96	440.53	407.10	373.67	340.24	306.81	273.37
139	620.80	583.82	548.06	514.41	481.96	450.20	418.44	386.67	354.91	323.15	291.38
136	624.79	594.43	565.29	538.25	512.43	487.28	462.14	436.99	411.85	386.70	361.56
133	626.06	595.53	566.22	539.02	513.02	487.71	462.40	437.08	411.77	386.46	361.14
131	626.90	596.38	567.08	539.87	513.88	488.57	463.27	437.96	412.65	387.34	362.03
129	627.75	597.23	567.93	540.73	514.74	489.43	464.13	438.82	413.51	388.21	362.90
126	629.02	598.53	569.26	542.09	516.13	490.86	465.58	440.31	415.03	389.76	364.48
122	630.66	600.21	570.98	543.85	517.94	492.71	467.47	442.24	417.00	391.77	366.54
115	633.53	603.48	574.65	547.93	522.41	497.58	472.74	447.91	423.08	398.24	373.41
113	634.35	604.50	575.86	549.33	524.01	499.37	474.73	450.09	425.45	400.81	376.17
106	637.22	607.72	579.44	553.26	528.29	504.01	479.72	455.43	431.15	406.86	382.57
97	640.32	611.30	583.49	557.78	533.29	509.48	485.67	461.86	438.05	414.24	390.43
91	642.38	613.56	585.95	560.45	536.16	512.55	488.94	465.32	441.71	418.10	394.49
86	643.84	615.19	587.74	562.41	538.28	514.84	491.39	467.95	444.51	421.06	397.62
77	646.47	618.04	590.81	565.69	541.79	518.56	495.33	472.11	448.88	425.66	402.43
71	648.23	619.72	592.43	567.25	543.28	519.98	496.69	473.40	450.11	426.82	403.53

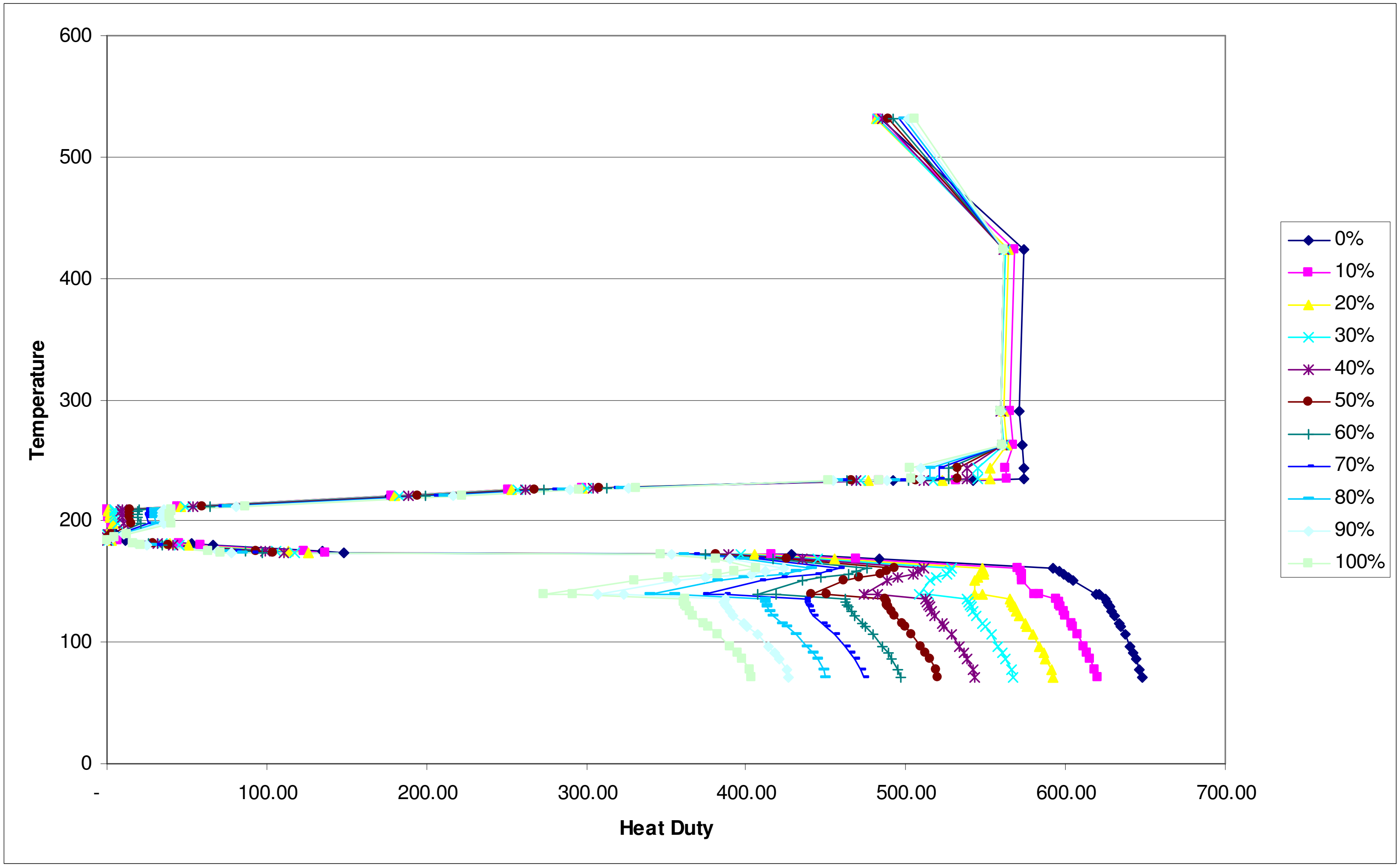


Figure C1 Grand composite curve for each step

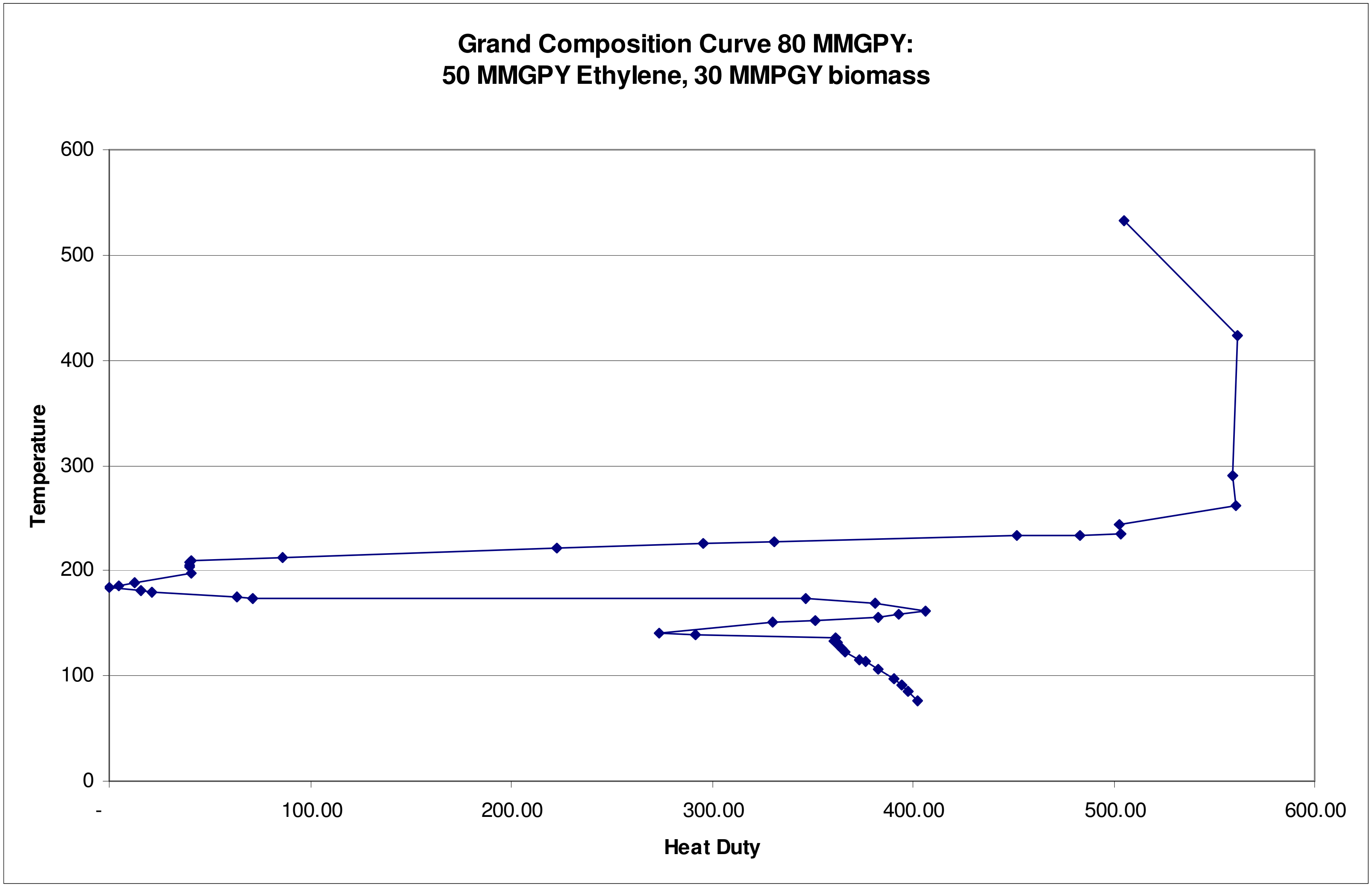


Figure C2 Grand composition curve for the optimal option

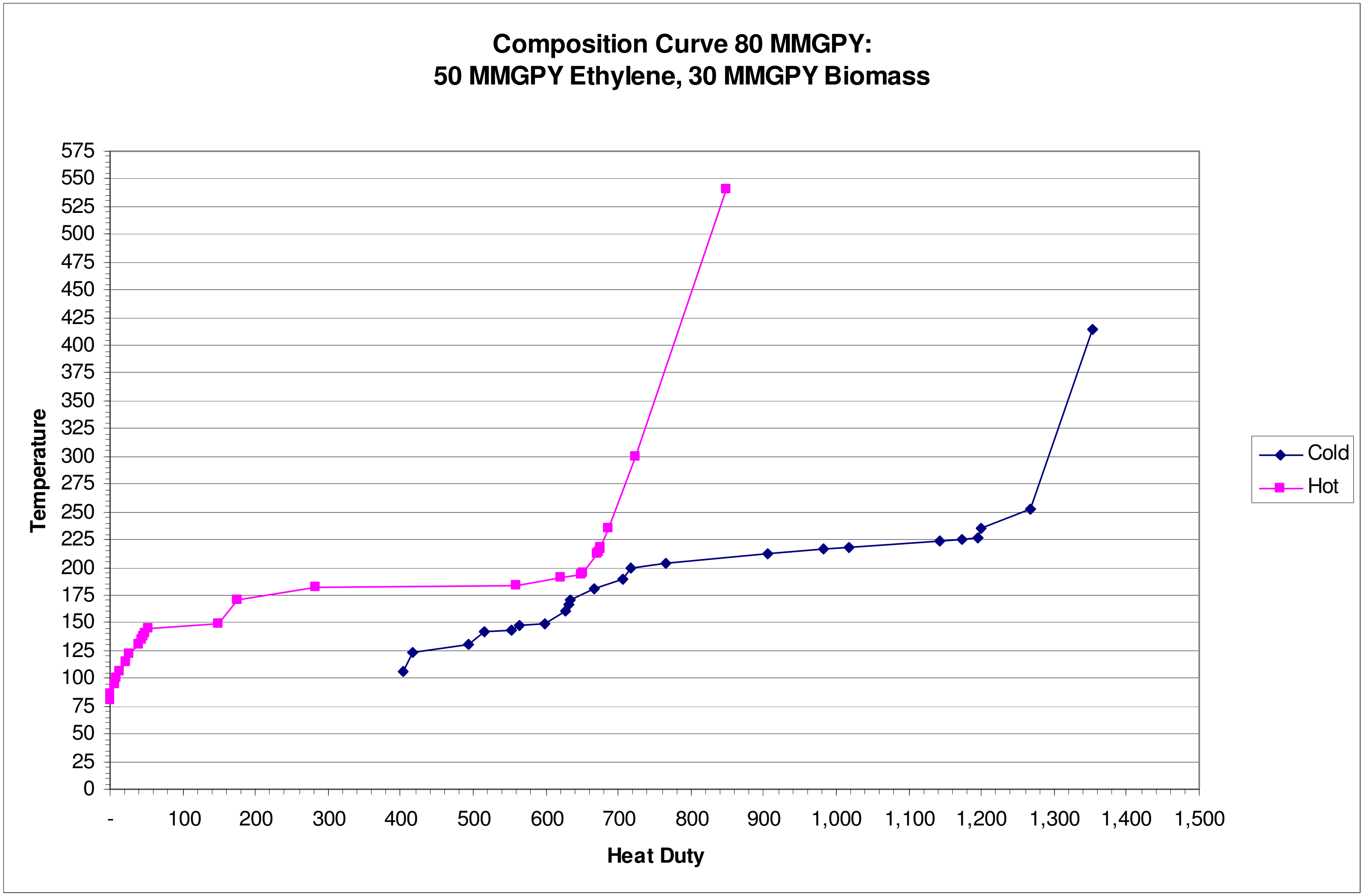


Figure C3 Composite curve for optimal option

APPENDIX D

Table D1 Rate of return calculation for each step for U.S. case

Year	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
0											
1	$ 7,796,538	$ 11,134,040	$ 12,640,266	$ 14,062,632	$ 15,435,522	$ 16,770,139	$18,069,938	$ 19,333,562	$ 20,553,611	$ 21,707,423	$ 22,590,373
2	$ 7,880,345	$ 10,765,653	$ 12,309,506	$ 13,697,957	$ 14,994,337	$ 16,219,700	$17,380,879	$ 18,475,892	$ 19,491,839	$ 20,388,447	$ 20,799,662
3	$ 6,833,512	$ 9,582,510	$ 11,023,737	$ 12,287,105	$ 13,444,341	$ 14,519,045	$15,518,644	$ 16,440,468	$ 17,269,168	$ 17,957,380	$ 18,076,281
4	$ 5,936,695	$ 8,539,316	$ 9,883,874	$ 11,034,343	$ 12,068,219	$ 13,010,998	$13,870,680	$ 14,644,294	$ 15,315,025	$ 15,831,022	$ 15,722,895
5	$ 5,169,364	$ 7,620,077	$ 8,874,105	$ 9,922,785	$ 10,847,317	$ 11,674,577	$12,413,112	$ 13,059,944	$ 13,597,553	$ 13,971,617	$ 13,689,369
6	$ 4,506,975	$ 6,805,196	$ 7,973,896	$ 8,930,267	$ 9,757,465	$ 10,483,286	$11,116,768	$ 11,655,146	$ 12,080,778	$ 12,338,469	$ 11,925,755
7	$ 3,971,088	$ 6,111,962	$ 7,206,195	$ 8,082,458	$ 8,825,327	$ 9,463,493	$10,006,594	$ 10,452,285	$ 10,783,191	$ 10,944,155	$ 10,430,378
8	$ 3,532,917	$ 5,517,558	$ 6,546,304	$ 7,352,568	$ 8,021,911	$ 8,583,887	$ 9,048,819	$ 9,414,944	$ 9,665,453	$ 9,745,934	$ 9,154,969
9	$ 3,143,094	$ 4,980,961	$ 5,946,840	$ 6,688,591	$ 7,291,635	$ 7,786,037	$ 8,182,717	$ 8,480,554	$ 8,663,575	$ 8,678,900	$ 8,035,516
10	$ 2,796,284	$ 4,496,549	$ 5,402,271	$ 6,084,575	$ 6,627,839	$ 7,062,346	$ 7,399,513	$ 7,638,898	$ 7,765,547	$ 7,728,691	$ 7,052,947
11	$ 2,487,741	$ 4,059,248	$ 4,907,570	$ 5,535,104	$ 6,024,472	$ 6,405,920	$ 6,691,273	$ 6,880,773	$ 6,960,605	$ 6,882,515	$ 6,190,525
12	$ 2,213,243	$ 3,664,476	$ 4,458,170	$ 5,035,254	$ 5,476,033	$ 5,810,507	$ 6,050,822	$ 6,197,888	$ 6,239,099	$ 6,128,983	$ 5,433,558
13	$ 1,969,033	$ 3,308,096	$ 4,049,923	$ 4,580,543	$ 4,977,521	$ 5,270,436	$ 5,471,671	$ 5,582,776	$ 5,592,382	$ 5,457,951	$ 4,769,152
14	$ 1,751,770	$ 2,986,375	$ 3,679,060	$ 4,166,895	$ 4,524,391	$ 4,780,563	$ 4,947,953	$ 5,028,712	$ 5,012,700	$ 4,860,387	$ 4,185,989
15	$ 1,558,479	$ 2,695,942	$ 3,342,158	$ 3,790,601	$ 4,112,512	$ 4,336,223	$ 4,474,363	$ 4,529,635	$ 4,493,106	$ 4,328,247	$ 3,674,133
16	$ 1,268,122	$ 2,323,837	$ 2,897,234	$ 3,293,231	$ 3,574,854	$ 3,767,365	$ 3,882,155	$ 3,921,808	$ 3,878,537	$ 3,719,935	$ 3,124,018
17	$ 1,019,235	$ 1,995,187	$ 2,501,420	$ 2,849,914	$ 3,095,889	$ 3,261,607	$ 3,357,209	$ 3,385,097	$ 3,338,499	$ 3,188,814	$ 2,650,450
18	$ 906,772	$ 1,801,150	$ 2,272,357	$ 2,592,551	$ 2,814,054	$ 2,958,449	$ 3,035,876	$ 3,049,142	$ 2,992,445	$ 2,839,687	$ 2,326,357
19	$ 806,719	$ 1,625,984	$ 2,064,271	$ 2,358,430	$ 2,557,876	$ 2,683,469	$ 2,745,299	$ 2,746,528	$ 2,682,261	$ 2,528,784	$ 2,041,894
20	$ 717,705	$ 1,467,852	$ 1,875,240	$ 2,145,451	$ 2,325,019	$ 2,434,048	$ 2,482,534	$ 2,473,948	$ 2,404,230	$ 2,251,920	$ 1,792,215

ROR	**12.40%**	**10.77%**	**10.08%**	**9.93%**	**10.02%**	**10.25%**	**10.58%**	**11.02%**	**11.56%**	**12.29%**	**13.93%**

Table D2 Rate of return calculation for each step for utilities case

Year	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
0											
1	$16,391,433	$ 18,746,836	$19,920,801	$ 20,890,046	$21,753,095	$22,544,702	$23,279,279	$ 23,960,641	$24,582,455	$ 25,104,979	$ 25,279,145
2	$12,919,050	$ 16,284,069	$17,794,396	$ 18,969,291	$19,966,230	$20,841,322	$21,617,758	$ 22,301,160	$22,879,801	$ 23,282,504	$ 23,078,607
3	$ 9,551,803	$ 12,956,478	$14,439,947	$ 15,570,978	$16,514,363	$17,328,969	$18,039,035	$ 18,650,181	$19,149,475	$ 19,456,244	$ 19,118,268
4	$ 7,068,794	$ 10,321,453	$11,733,232	$ 12,798,928	$13,678,386	$14,428,993	$15,074,259	$ 15,619,210	$16,050,120	$ 16,281,556	$ 15,858,526
5	$ 5,236,903	$ 8,234,145	$ 9,548,704	$ 10,537,461	$11,348,312	$12,034,664	$12,618,240	$ 13,103,148	$13,475,209	$ 13,647,611	$ 13,175,254
6	$ 3,882,212	$ 6,574,437	$ 7,777,889	$ 8,683,708	$ 9,424,204	$10,047,429	$10,572,741	$ 11,003,187	$11,324,432	$ 11,450,767	$ 10,955,939
7	$ 2,890,543	$ 5,280,880	$ 6,377,255	$ 7,205,727	$ 7,882,347	$ 8,449,486	$ 8,923,984	$ 9,307,781	$ 9,586,452	$ 9,676,363	$ 9,171,291
8	$ 2,160,678	$ 4,264,984	$ 5,260,138	$ 6,016,988	$ 6,635,633	$ 7,152,772	$ 7,582,714	$ 7,926,304	$ 8,169,092	$ 8,230,147	$ 7,723,879
9	$ 1,615,104	$ 3,444,519	$ 4,338,708	$ 5,024,357	$ 5,586,105	$ 6,055,060	$ 6,443,036	$ 6,749,868	$ 6,961,289	$ 7,000,081	$ 6,504,897
10	$ 1,207,289	$ 2,781,888	$ 3,578,688	$ 4,195,481	$ 4,702,577	$ 5,125,810	$ 5,474,651	$ 5,748,041	$ 5,932,061	$ 5,953,859	$ 5,478,295
11	$ 902,447	$ 2,246,730	$ 2,951,802	$ 3,503,347	$ 3,958,792	$ 4,339,169	$ 4,651,813	$ 4,894,907	$ 5,055,004	$ 5,064,004	$ 4,613,711
12	$ 674,578	$ 1,814,521	$ 2,434,728	$ 2,925,395	$ 3,332,648	$ 3,673,251	$ 3,952,648	$ 4,168,396	$ 4,307,620	$ 4,307,145	$ 3,885,575
13	$ 504,246	$ 1,465,457	$ 2,008,232	$ 2,442,788	$ 2,805,538	$ 3,109,530	$ 3,358,567	$ 3,549,716	$ 3,670,737	$ 3,663,405	$ 3,272,354
14	$ 376,924	$ 1,183,543	$ 1,656,446	$ 2,039,798	$ 2,361,799	$ 2,632,321	$ 2,853,776	$ 3,022,861	$ 3,128,017	$ 3,115,878	$ 2,755,912
15	$ 281,750	$ 955,862	$ 1,366,283	$ 1,703,289	$ 1,988,244	$ 2,228,347	$ 2,424,855	$ 2,574,202	$ 2,665,539	$ 2,650,183	$ 2,320,974
16	$ 203,305	$ 734,962	$ 1,067,680	$ 1,343,515	$ 1,578,092	$ 1,776,509	$ 1,939,313	$ 2,063,206	$ 2,138,886	$ 2,125,145	$ 1,850,574
17	$ 146,323	$ 562,648	$ 830,081	$ 1,053,821	$ 1,245,168	$ 1,407,665	$ 1,541,400	$ 1,643,402	$ 1,705,802	$ 1,694,068	$ 1,467,818
18	$ 109,376	$ 454,410	$ 684,674	$ 879,971	$ 1,048,226	$ 1,191,636	$ 1,309,728	$ 1,399,485	$ 1,453,599	$ 1,440,875	$ 1,236,167
19	$ 81,758	$ 366,994	$ 564,738	$ 734,801	$ 882,433	$ 1,008,759	$ 1,112,877	$ 1,191,771	$ 1,238,684	$ 1,225,524	$ 1,041,076
20	$ 61,114	$ 296,394	$ 465,812	$ 613,580	$ 742,862	$ 853,948	$ 945,612	$ 1,014,887	$ 1,055,544	$ 1,042,359	$ 876,773
ROR	**33.78%**	**23.82%**	**21.24%**	**19.76%**	**18.79%**	**18.13%**	**17.69%**	**17.43%**	**17.35%**	**17.57%**	**18.74%**

Table D3 Rate of return calculation for each step for Middle East case

Year	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
0											
1	$ 31,444,842	$36,863,213	$38,194,966	$38,699,358	$38,753,900	$38,508,332	$38,040,046	$ 37,394,112	$36,595,876	$ 35,646,172	$34,281,950
2	$ 16,724,811	$23,868,509	$26,670,061	$28,548,998	$29,910,491	$30,918,274	$31,658,377	$ 32,181,343	$32,514,082	$ 32,646,304	$32,093,515
3	$ 8,687,253	$14,913,907	$17,856,601	$20,084,209	$21,902,955	$23,438,554	$24,758,727	$ 25,904,840	$26,899,872	$ 27,731,500	$27,857,176
4	$ 4,514,049	$ 9,323,871	$11,963,319	$14,139,356	$16,051,763	$17,783,419	$19,380,437	$ 20,872,710	$22,277,894	$ 23,581,906	$24,206,104
5	$ 2,346,569	$ 5,832,738	$ 8,020,879	$ 9,962,327	$11,774,284	$13,505,947	$15,186,457	$ 16,837,031	$18,472,102	$ 20,078,376	$21,060,035
6	$ 1,220,140	$ 3,650,141	$ 5,379,974	$ 7,022,673	$ 8,641,265	$10,263,256	$11,907,434	$ 13,590,604	$15,327,164	$ 17,107,879	$18,336,332
7	$ 635,457	$ 2,289,814	$ 3,618,943	$ 4,966,499	$ 6,364,693	$ 7,829,746	$ 9,376,135	$ 11,020,325	$12,779,763	$ 14,652,023	$16,047,597
8	$ 331,428	$ 1,439,582	$ 2,440,623	$ 3,522,600	$ 4,703,053	$ 5,994,383	$ 7,411,273	$ 8,973,044	$10,702,745	$ 12,607,208	$14,110,321
9	$ 172,859	$ 905,050	$ 1,645,961	$ 2,498,482	$ 3,475,220	$ 4,589,245	$ 5,858,168	$ 7,306,093	$ 8,963,293	$ 10,847,763	$12,406,914
10	$ 90,156	$ 568,996	$ 1,110,040	$ 1,772,104	$ 2,567,939	$ 3,513,485	$ 4,630,531	$ 5,948,816	$ 7,506,544	$ 9,333,865	$10,909,143
11	$ 47,022	$ 357,722	$ 748,613	$ 1,256,904	$ 1,897,523	$ 2,689,892	$ 3,660,157	$ 4,843,686	$ 6,286,551	$ 8,031,243	$ 9,592,184
12	$ 24,525	$ 224,896	$ 504,866	$ 891,487	$ 1,402,134	$ 2,059,357	$ 2,893,135	$ 3,943,858	$ 5,264,836	$ 6,910,414	$ 8,434,210
13	$ 12,791	$ 141,390	$ 340,483	$ 632,307	$ 1,036,077	$ 1,576,625	$ 2,286,850	$ 3,211,195	$ 4,409,174	$ 5,946,006	$ 7,416,026
14	$ 6,671	$ 88,890	$ 229,622	$ 448,478	$ 765,587	$ 1,207,050	$ 1,807,618	$ 2,614,641	$ 3,692,578	$ 5,116,190	$ 6,520,759
15	$ 3,479	$ 55,884	$ 154,858	$ 318,093	$ 565,714	$ 924,107	$ 1,428,814	$ 2,128,911	$ 3,092,446	$ 4,402,182	$ 5,733,569
16	$ 1,792	$ 34,461	$ 102,072	$ 219,831	$ 406,142	$ 685,484	$ 1,091,307	$ 1,670,511	$ 2,489,459	$ 3,632,645	$ 4,833,908
17	$ 922	$ 21,228	$ 67,188	$ 151,676	$ 291,028	$ 507,374	$ 831,472	$ 1,307,190	$ 1,997,889	$ 2,987,555	$ 4,061,613
18	$ 481	$ 13,346	$ 45,312	$ 107,580	$ 215,049	$ 388,441	$ 657,229	$ 1,064,349	$ 1,673,185	$ 2,570,616	$ 3,571,293
19	$ 251	$ 8,390	$ 30,558	$ 76,303	$ 158,906	$ 297,387	$ 519,500	$ 866,622	$ 1,401,252	$ 2,211,865	$ 3,140,164
20	$ 131	$ 5,275	$ 20,609	$ 54,120	$ 117,420	$ 227,677	$ 410,634	$ 705,627	$ 1,173,515	$ 1,903,180	$ 2,761,081
ROR	**91.73%**	**59.06%**	**48.28%**	**40.99%**	**35.33%**	**30.62%**	**26.51%**	**22.82%**	**19.41%**	**16.22%**	**13.73%**

APPENDIX E

Table E1 Heat duty data from linear interpolation HEN

	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
MMBtu/hr	648	620	592	567	543	520	497	473	450	427	403
lb/yr											
Cost $	$ 38,115,924	$ 36,439,536	$ 34,834,884	$ 33,354,300	$ 31,944,864	$ 30,574,824	$ 29,205,372	$ 27,835,920	$ 26,466,468	$ 25,097,016	$ 668,764
Area ft^2	216,077	206,573	197,477	189,083	181,093	173,327	165,563	157,800	150,037	142,273	134,177
Cost $	$ 3,175,373	$ 3,090,826	$ 3,008,427	$ 2,931,041	$ 2,856,087	$ 2,781,949	$ 2,706,503	$ 2,629,627	$ 2,551,223	$ 2,471,178	$ 2,385,812
MMBtu/hr	484	482	481	483	485	489	492	495	499	502	505
lb/yr											
Cost $	$ 24,394,104	$ 24,298,344	$ 24,264,072	$ 24,335,640	$ 24,468,192	$ 24,635,016	$ 24,801,840	$ 24,969,168	$ 25,135,992	$ 25,302,816	$ 25,469,640
Area ft^2	121,003	120,528	120,358	120,713	121,370	122,198	123,025	123,855	124,683	125,510	126,338
Cost $	$ 2,242,367	$ 2,237,081	$ 2,235,187	$ 2,239,140	$ 2,246,450	$ 2,255,628	$ 2,264,780	$ 2,273,935	$ 2,283,039	$ 2,292,118	$ 2,301,173

	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
TCI	$ 5,417,739	$ 5,327,907	$ 5,243,614	$ 5,170,182	$ 5,102,538	$ 5,037,577	$ 4,971,283	$ 4,903,563	$ 4,834,262	$ 4,763,297	$ 4,686,985
Operating Cost	$ 62,510,028	$ 60,737,880	$59,098,956	$ 57,689,940	$ 56,413,056	$ 55,209,840	$ 54,007,212	$ 52,805,088	$ 51,602,460	$ 50,399,832	$ 49,138,404

Table E2 Heat duty data from calculated HEN

	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1.0
MMBtu/hr	648	624	599	575	550	525	501	476	452	427	403
lb/yr											
Cost $	$ 38,115,924	$36,671,208	$ 35,226,492	$33,781,776	$ 32,337,060	$ 30,892,344	$29,447,628	$ 28,002,912	$ 26,558,196	$25,113,480	$23,668,764
Area ft^2	216,077	207,887	199,697	191,507	183,317	175,127	166,937	158,747	150,557	142,367	134,177
Cost $	$ 3,175,373	$ 3,102,601	$ 3,028,674	$ 2,953,523	$ 2,877,075	$ 2,799,248	$ 2,719,951	$ 2,639,081	$ 2,556,525	$ 2,472,151	$ 2,385,812
MMBtu/hr	484	486	488	490	493	495	497	499	501	503	505
lb/yr											
Cost $	$ 24,394,104	$24,501,658	$ 24,609,211	$24,716,765	$ 24,824,318	$ 24,931,872	$25,039,426	$ 25,146,979	$ 25,254,533	$25,362,086	$25,469,640
Area ft^2	121,003	121,536	122,070	122,603	123,137	123,670	124,204	124,737	125,271	125,804	126,338
Cost $	$ 2,242,367	$ 2,248,293	$ 2,254,210	$ 2,260,116	$ 2,266,011	$ 2,271,897	$ 2,277,772	$ 2,283,638	$ 2,289,493	$ 2,295,338	$ 2,301,173

	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
TCI	$ 5,417,739	$ 5,350,895	$ 5,282,883	$ 5,213,638	$ 5,143,086	$ 5,071,145	$ 4,997,723	$ 4,922,719	$ 4,846,017	$ 4,767,489	$ 4,686,985
Operating Cost	$ 62,510,028	$61,172,866	$ 59,835,703	$58,498,541	$ 57,161,378	$ 55,824,216	$54,487,054	$ 53,149,891	$ 51,812,729	$50,475,566	$49,138,404

Table E3 Saving between linear interpolation and calculated HEN

	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1
Revenue saving		$ 260,991	$ 442,048	$ 485,160	$ 448,993	$ 368,626	$ 287,905	$ 206,882	$ 126,161	$ 45,441	
Operaring Cost saving		$ 434,986	$ 736,747	$ 808,601	$ 748,322	$ 614,376	$ 479,842	$ 344,803	$ 210,269	$ 75,734	
TCI Saving		$ 22,988	$ 39,269	$ 43,457	$ 40,549	$ 33,568	$ 26,440	$ 19,156	$ 11,756	$ 4,193	
Increment TCI	0.00%	0.43%	0.74%	0.83%	0.79%	0.66%	0.53%	0.39%	0.24%	0.09%	0.00%
Increment OC	0.00%	0.71%	1.23%	1.38%	1.31%	1.10%	0.88%	0.65%	0.41%	0.15%	0.00%

APPENDIX F

Table F1 Aspen Plus NREL properties (1of2)

Property	Aspen	Units	Glucose	Xylose	Cellulose	Xylan	Lignin	Cellulase	Zymo	Biomass	Solslds	Solunkn	Gypsum
Molecular weight	MW		180.16	150.132	162.14	132.117	122.493	22.83	24.62	23.238	16.5844	15.0134	172.168
Critical temperature	TC	K	1011.1	890.42							1011.1	1011.1	
Critical pressure	PC	Pascal	6,200,000.0	6,577,700.0							6,200,000.0	6,200,000.0	
Critical volume	VC	cum/Kmole	0.4165	0.3425							0.4165	0.4165	
Acentric factor	OMEGA		2.5674	2.3042							2.5674	2.5674	
I.G. heat of formation	DHFORM	J/kmole	-1,256,903,000.0	-1,040,020,000.0							-47,540,000.0	-119,000,000.0	
I.G. free energy of form.	DGFORM	J/kmole	-909,330,000.0										
Solid heat of formation	DHSFRM	J/kmole			-976,262,000.0	-762,416,000.0	-1,592,659,000.0	-74,944,000.0	-130,500,000.0	-97,133,800.0			-2,022,628,000.0
Solid free energy of form.	DGSFRM	J/kmole											-1,797,197,000.0
Vapor pressure	PLXANT/1	Pascal	1182.2	481.33							1182.2	1182.2	
	PLXANT/2		-84682	-46623							-84682	-84682	
	PLXANT/3		0	0							0	0	
	PLXANT/4		0.1564	2.10E-02							0.1564	0.1564	
	PLXANT/5		-175.85	64.331							-175.85	-175.85	
	PLXANT/6		-2.38E-05	6.22E-06							-2.38E-05	-2.38E-05	
	PLXANT/7		2	2							2	2	
	PLXANT/8		573.15	573.15							573.15	573.15	
	PLXANT/9		993.15	993.15							993.15	993.15	
Heat of vaporization	DHVLWT/1	J/kmole	502,000.0	4,186,800.0							4,186,800.0	4,186,800.0	
	DHVLWT/2		298	298							298	298	
	DHVLWT/3		0.38	0.38							0.38	0.38	
	DHVLWT/4		0	0							0	0	
	DHVLWT/5		200	200							200	200	
Liquid molar volume	RKTZRA	cum/Kmole	0.35852	0.29936									
Solid molar volume	VSPOLY/1	cum/Kmole			0.106	0.0884	0.0817	0.0152	0.0164	0.01549			0.0746
	VSPOLY/2				0	0	0	0	0	0			0
	VSPOLY/3				0	0	0	0	0	0			0
	VSPOLY/4				0	0	0	0	0	0			0
	VSPOLY/5				0	0	0	0	0	0			0
	VSPOLY/6				298.15	298.15	298.15	298.15	298.15	298.15			298.15
	VSPOLY/7				1000	1000	1000	1000	1000	1000			1000

Table F2 Aspen Plus NREL properties (2of2)

I.G. Heat capacity		CPIG/1	J/kmole K	-5846.2	-4349.1							-5846.2	-5846.2	
		CPIG/2		1,005.4	832.4							1,005.4	1,005.4	
		CPIG/3		-0.85893	-0.707							-0.85893	-0.85893	
		CPIG/4		2.87E-04	2.34E-04							2.87E-04	2.87E-04	
		CPIG/5		-5.65E-10	-2.03E-10							-5.65E-10	-5.65E-10	
		CPIG/6		0	0							0	0	
		CPIG/7		573.15	573.15							573.15	573.15	
		CPIG/8		1,033.2	1,033.2							1,033.2	1,033.2	
		CPIG/9		0	0							0	0	
		CPIG/10		0	0							0	0	
		CPIG/11		0	0							0	0	
Solid heat caoacity		CPSPO1/1	J/kmole K			-11704	-9529.9	31431.7	35533	38409	35910			72182
		CPSPO1/2				672.07	547.25	394.42	0	0	0			97.34
		CPSPO1/3				0	0	0	0	0	0			0
		CPSPO1/4				0	0	0	0	0	0			0
		CPSPO1/5				0	0	0	0	0	0			- 137,330,000.0
		CPSPO1/6				0	0	0	0	0	0			0
		CPSPO1/7				298.15	298.15	298.15	298.15	298.15	298.15			298
		CPSPO1/8				1000	1000	1000	1000	1000	1000			1400
Liquid heat capacity		CPLDIP/1	J/kmole K	207,431.0	172,857.0							19094	17286	
		CPLDIP/2		0	0							0	0	
		CPLDIP/3		0	0							0	0	
		CPLDIP/4		0	0							0	0	
		CPLDIP/5		0	0							0	0	
		CPLDIP/6		250	250							250	250	
		CPLDIP/7		1000	1000							1000	1000	

VITA

Benjamin R. Cormier was on born in 1978, in Chambray-les-Tours, France. He grew up in a small town called Civry located one hour south of Paris. After the middle school, he went to a boarding school to find new challenges. Then, he obtained his baccalaureate in science from Saint Marie at Blois, France in 1997. Next, he attended Francois Rabelais University (in Tours, France) for two years and graduated with a diploma in science combining mathematics, physics, and chemistry. With the support of his grandparents, Claude and Lucien Monier, he enrolled in a study-abroad program which brought him to the United States in 1999 for two years. After that, he transferred to Virginia Tech where he obtained a bachelor in chemical engineering in 2004. During this time at Virginia Tech, he found an interest in organic chemistry, and he then was hired in a research lab. At the same time, his passion for chemical engineering grew and he finally decided to join Dr S. Ted Oyama's research lab working on environmental catalysis. During his last year at Virginia Tech, he was encouraged to apply for graduate school by Dr Y.A. Liu and Dr S. Ted Oyama. He entered the graduate program at Texas A&M University in September, 2004 and graduated with a Master of Science in Chemical Engineering in 2005.

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